

PSDF

Power Systems Development Facility

Topical Report

Gasification Test Run TC15

*April 19, 2004 -
April 29, 2004*

*DOE Cooperative Agreement Number
DE-FC21-90MC25140*

**SOUTHERN
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POWER SYSTEMS DEVELOPMENT FACILITY TOPICAL REPORT

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POWER SYSTEMS DEVELOPMENT FACILITY

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ABSTRACT

This report discusses Test Campaign TC15 of the Kellogg Brown & Root, Inc. (KBR) Transport Gasifier train with a Siemens Power Generation, Inc. (SPG) particle filter system at the Power Systems Development Facility (PSDF) located in Wilsonville, Alabama. The Transport Gasifier is an advanced circulating fluidized-bed reactor designed to operate as either a combustor or gasifier using a particulate control device (PCD). While operating as a gasifier, either air or oxygen can be used as the oxidant.

Test run TC15 began on April 19, 2004, with the startup of the main air compressor and the lighting of the gasifier startup burner. The Transport Gasifier was shutdown on April 29, 2004, accumulating 200 hours of operation using Powder River Basin (PRB) subbituminous coal. About 91 hours of the test run occurred during oxygen-blown operations. Another 6 hours of the test run was in enriched-air mode. The remainder of the test run, approximately 103 hours, took place during air-blown operations. The highest operating temperature in the gasifier mixing zone mostly varied from 1,800 to 1,850°F. The gasifier exit pressure ran between 200 and 230 psig during air-blown operations and between 110 and 150 psig in oxygen-enhanced air operations.

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1.0 EXECUTIVE SUMMARY

1.1 SUMMARY

This report discusses test campaign TC15 of the Kellogg Brown & Root, Inc. (KBR) Transport Gasifier train with a Siemens Power Generation, Inc. (SPG) particulate control device (PCD) at the Power Systems Development Facility (PSDF) located near Wilsonville, Alabama. The PSDF is a flexible test facility designed to develop advanced coal-fired power system components and assess the associated integration and control issues. TC15 began on April 19, 2004, and lasted until April 29, 2004, accumulating 200 hours of on-coal operation.

1.2 TEST CAMPAIGN OBJECTIVES

TC15 evaluated gasifier and PCD operations with Powder River Basin (PRB) subbituminous coal using air and pure oxygen as the gasification oxidants. The primary test objectives were:

- Syngas Cleanup Commissioning.
- Oxygen Distribution System Commissioning.
- Failsafe Device Testing.
- Higher Face Velocity Testing.

1.3 TEST CAMPAIGN SUMMARY

On April 19, 2004, coal feed began at a low rate to heat the gasifier to 1,700°F. Once the gasifier was at operating temperature, commissioning of the hot gas cleanup system began. During this time, testing on the developmental feeder occurred. The feeder ran well for short periods of time, feeding 3,000 to 5,000 pph of coal, but eventually fine coal particles began to pack in the lock hopper, causing the feeder to run out of coal a few times. When maintaining a constant coal-feed rate became difficult, the developmental feeder testing was aborted to avoid causing a gasifier upset.

Pilot syngas burner (PSB) testing began on April 22, 2004, using syngas from the Transport Gasifier and enrichment propane to achieve a mixture gas with a lower heating value of about 90 Btu/scf. The gasifier temperature controls were tuned at this time also. When PSB testing concluded, the gasifier pressure was lowered in preparation for oxygen-blown mode.

Oxygen-blown operations began on April 23, 2004, and gas cleanup testing continued with the richer syngas. A restriction formed around the new oxygen nozzles in the upper mixing zone (UMZ) due to a temperature excursion. Initially the oxygen flow rate was excessive, since the associated flow meter was not reading correctly. During later oxygen-blown operations, PCD failsafe testing began as did operations with a lower PCD back-pulse pressure. Neither adversely affected the gasifier system.

On April 25, 2004, standpipe operations were unstable, adversely impacting circulation in the gasifier. During the attempt to stabilize gasifier conditions, a restriction in the gasifier J-leg stopped solids circulation. Efforts to remove the restriction were partially successful, but the circulation in the gasifier remained low, limiting the system performance. After other attempts to restore gasifier conditions proved unsuccessful, the test run ended on April 29, 2004.

1.4 TEST CAMPAIGN PERFORMANCE

Performance of the major equipment during TC15 is summarized in the following three sections:

1.4.1 Transport Gasifier Performance

- During air-blown gasification, the raw lower heating values at the exit of the gasifier were between 26 and 74 Btu/scf, resulting in projected heating values at the turbine inlet of a commercial gasifier of between 106 and 135 Btu/scf. The projection converts PSDF data into commercial projections by accounting for the use of recycled gas, the lower heat loss per pound of coal fed, and the use of cold gas cleanup in a commercial gasifier. The heating values were slightly lower in TC15 than in previous test runs due to a lower circulation rate and coal-feed rate.
- In oxygen-blown gasification, the raw lower heating values at the exit of the gasifier ranged from 80 to 98 Btu/scf, resulting in projected heating values of 256 to 275 Btu/scf.
- The carbon conversion was between 91 and 98 percent for both air- and oxygen-blown gasification. The TC15 carbon conversions were typical for PRB coal.
- The raw cold gas efficiency ranged from 36 to 61 percent during air-blown operations. It was between 60.6 and 64.3 percent for the oxygen-blown operating periods. The commercially projected efficiency was between 68 and 81 percent, based on data from the air-blown operating periods. The commercially projected efficiency was between 83 and 91 percent, based on data from the oxygen-blown operating periods.
- The hot gas efficiency ranged from 70.4 to 86.6 percent for both the air- and oxygen-blown periods.
- The sulfur concentration of the syngas at the gasifier exit ranged from 268 to 410 ppm for air-blown gasification and 574 to 709 for oxygen-blown gasification. No limestone was injected into the gasifier during the test campaign.
- In the middle of the test run, a deposit formed around the two new oxygen nozzles and in the lower riser during the commissioning of the new oxygen system. Initially, the oxygen flow was excessive, since the flow indication was not reading properly at the time. The high flow rate caused a temperature excursion in the gasifier, creating the deposit. Parts of the deposit worked their way to the gasifier J-leg where they hindered circulation. Although adjustments in aerations flows and gasifier pressure were able to dislodge the material, the deposit in the riser continued to restrict circulation and cause an uneven temperature profile in the gasifier.

1.4.2 PCD Performance

- PCD operation was stable throughout TC15. There were no filter element failures or gasification ash bridging in the PCD.
- Outlet loading samples indicated good sealing of the filter vessel, and the outlet loading was maintained below the detection limit of 0.1 ppmw for most of the test run.

- As in TC14, high face velocity was tested by leaving several filter element slots blank. The pressure drop and back-pulse frequency requirements in TC15 were similar to those in TC14.
- Failsafe testing with gasification ash injection was performed on the CeraMem ceramic failsafe. The effect of injection was immediate and the failsafe seemed to plug quickly. The first outlet sample showed approximately 0.5 ppmw, and the second outlet sample, taken about 4 hours later, showed approximately 0.17 ppmw.
- The physical and chemical characteristics of the gasification ash were generally consistent with those from previous PRB tests, although the normalized drag was at the upper end of the range from previous tests. The TC15 data continued to show that the performance of the PCD is affected by the carbon conversion achieved in the gasifier. As carbon conversion drops, drag increases, and solids carryover to the PCD increases, resulting in increased pressure drop across the PCD.

1.4.3 Performance of Other Systems

- The hot gas cleanup unit was commissioned in TC15 for a period of 35 hours using two desulfurizer sorbents, RVS-1 and RVSLT-1. RVS-1 achieved a 19-percent absorption capacity, while RVSLT-1 achieved a 30-percent absorption capacity. Condensation in the process lines and in the gas sample lines was the major operational challenge.
- The new minireactor was commissioned while passing a mixture of bottled gases through a nickel-based ammonia cracking catalyst. The test unit proved to be sufficiently flexible for testing the catalyst performance. The nickel catalyst was effective in decomposing ammonia when the minireactor temperature exceeded 1,500°F. At the pressure tested, the catalyst converted about 98 percent of the ammonia. The hydrogen and nitrogen concentrations and system pressure proved to be important factors in controlling the exit ammonia concentration.
- The PSB operated on syngas for around 15 hours at syngas flow rates up to 17,000 pph. The unit produced 23.5 MWh. At that time, the wet raw syngas heating value was approximately 60 Btu/scf. As the propane flow rate decreased, the flame remained stable to the point where about 86 percent of the total energy input to the combustion turbine (CT) came from syngas. Both the PSB wall temperatures and the combustor noise remained low throughout the run.

1.5 CONCLUSIONS

- The Transport Gasifier operated for 200 hours in TC15 using PRB coal. Approximately 91 of the hours were in oxygen-blown gasification and the balance was in air-blown gasification and air-to-oxygen transitions.
- The gasifier experienced stable operations during the air-blown portion of the test run. Oxygen-blown operations in the latter half of the test run were not smooth due to poor solids circulation. The low circulation rate caused the temperature profile to be nonuniform.
- In the middle of the test run, a deposit formed in the riser during the commissioning of the new oxygen nozzles in the UMZ. Parts of the deposit worked their way to the gasifier J-leg where they hindered circulation. Although adjustments in aerations flows and in the gasifier pressure were able to dislodge the material, the deposit in the riser continued to restrict circulation and cause uneven temperature profiles in the gasifier.
- During the previous test run, erosion in the primary cyclone caused the gasifier to continuously lose about 100 pph of bed material. In TC15, the repaired cyclone had a higher efficiency, and thus the solids carry-over rate was lower even at a high coal-feed rate.
- The test run ended earlier than scheduled on the afternoon of April 29, 2004. The shutdown was smooth, leaving no transient cake on the PCD filter candles. Initial post-run inspections showed a deposit at the transition between the riser and mixing zone, as well as, another smaller deposit in the gasifier J-leg.

2.0 OPERATING SUMMARY

2.1 PROJECT/SYSTEM DESCRIPTION

The PSDF, near Wilsonville, Alabama, is funded by the U.S. Department of Energy, Southern Company, and participants currently including the Electric Power Research Institute, SPG, KBR, Burlington Northern Santa Fe Corporation (BNSF), Peabody Energy, and the Lignite Energy Council. The PSDF is an engineering scale demonstration of key features of advanced coal-fired power systems designed at sufficient size to evaluate system components and assess the integration and control issues of these advanced power systems. The facility also supports clean coal technology programs to address environmental concerns associated with using fossil fuels for producing electricity, chemicals, and transportation fuels.

The KBR Transport Reactor, which operates at the PSDF, is a pressurized, advanced circulating fluidized bed reactor operating in either combustion or gasification mode. While operating in gasification mode, either air or oxygen can be used as the oxidant. The particulate-laden gas exiting the reactor is filtered by a downstream high temperature, high pressure filter vessel, the PCD. In gasification mode, the objective of the PCD is to clean the gas sufficiently so that it can be utilized in a downstream gas turbine/combustor or fuel cell. A gas cleanup skid is also available to remove various pollutants from a syngas slipstream. A flow diagram of the gasifier train is shown in [Figure 2.1-1](#). The Transport Reactor train has now operated for about 5,000 hours in combustion mode and over 5,300 hours during gasification.

The Transport Gasifier, shown in [Figure 2.1-2](#), consists of a mixing zone, a riser, a disengager, a cyclone, a standpipe, a loopseal, and a J-leg. Steam and either air or oxygen are mixed together and introduced in the lower mixing zone (LMZ) while the fuel, sorbent, and additional air and steam (if needed) are added in the UMZ. The steam and oxidant, along with the fuel, sorbent, and solids from the standpipe, are mixed together in the UMZ. The UMZ, located below the riser, has a slightly larger diameter than the riser. The gas and solids move up the riser before entering the disengager, which removes larger particles by gravity separation. The majority of the solids flow from the disengager into the standpipe, and the remaining solids flow, along with the syngas, to the cyclone, which removes most of the particles not collected by the disengager. The solids collected by the disengager and cyclone are recycled back to the gasifier mixing zone through the standpipe and a J-leg. The nominal gasifier operating temperature is 1,800°F, and the gasifier system is designed to have a maximum operating pressure of 294 psig with a thermal capacity of about 41 MBtu/hr. Due to a lower oxygen supply pressure, the maximum operating pressure is about 180 psi during oxygen-blown gasification.

For startup purposes, a direct propane-fired burner is provided at the gasifier mixing zone. Coal and sorbent (when required for sulfur capture) are separately fed into the Transport Gasifier through lockhoppers. Coal is ground to a nominal particle diameter between 250 and 400 μm . Sorbent, either limestone or dolomitic sorbent, is ground to a nominal particle diameter of 10 to 100 μm .

The gas exits the Transport Gasifier cyclone and goes to the primary gas cooler and then to the PCD for final particulate cleanup. The metal or ceramic filter elements used in the PCD remove almost all the dust from the gas stream, preventing erosion of downstream equipment and controlling particulate emissions from the plant. Shown in [Figure 2.1-3](#), the PCD utilizes a tube sheet holding up to 91 filter elements, which are attached to one of two plenums. Process gas flows into the PCD through a tangential entrance, around a shroud, and through the filter elements into the plenums. Failsafe devices are located on the clean side of the filter elements to stop solids leakage in the event of element failures. High pressure nitrogen back-pulsing, typically lasting 0.2 seconds, is used to clean the filters periodically to remove the accumulated solids and control the pressure drop across the tube sheet. The solids fall to the PCD cone and are removed through a lock hopper system or the continuous fine ash depressurization (CFAD) system.

After exiting the PCD, a portion of the syngas can flow to the PSB, where the gas is combusted using air from the turbine compressor. Propane supplied to the PSB serves as a pilot for the burner, as well as, a supplement to the syngas fuel to maintain burner flame stability. After combusting in the burner, the gas passes through the turbine before exiting the turbine stack. An associated generator supplies power to the electricity transmission grid. The PSB and turbine system are capable of running independently of the gasifier by using propane alone as fuel. A small portion of the syngas can also flow to a specialized gas cleanup system downstream of the PCD. The gas cleanup system removes sulfur, nitrogen, and chlorine compounds, providing a syngas suitable for use in a fuel cell.

The syngas not flowing to the PSB or cleanup system continues to the secondary gas cooler and then passes through a pressure control valve. The gas is then sent to the atmospheric syngas combustor (thermal oxidizer) which oxidizes carbon monoxide, hydrogen, methane, other organics, reduced sulfur compounds (H_2S , COS , and CS_2), and reduced nitrogen compounds (NH_3 and HCN). The atmospheric syngas combustor uses propane as a supplemental fuel. The gas from the atmospheric syngas combustor goes to the heat recovery boiler, through the baghouse, and then to the stack.

The Transport Gasifier produces fine solids filtered by the PCD and coarse solids extracted from the gasifier standpipe. The two solid streams are cooled using screw coolers, reduced in pressure in lock hoppers and then combined together. The CFAD system is also available for removing fine solids from the PCD.

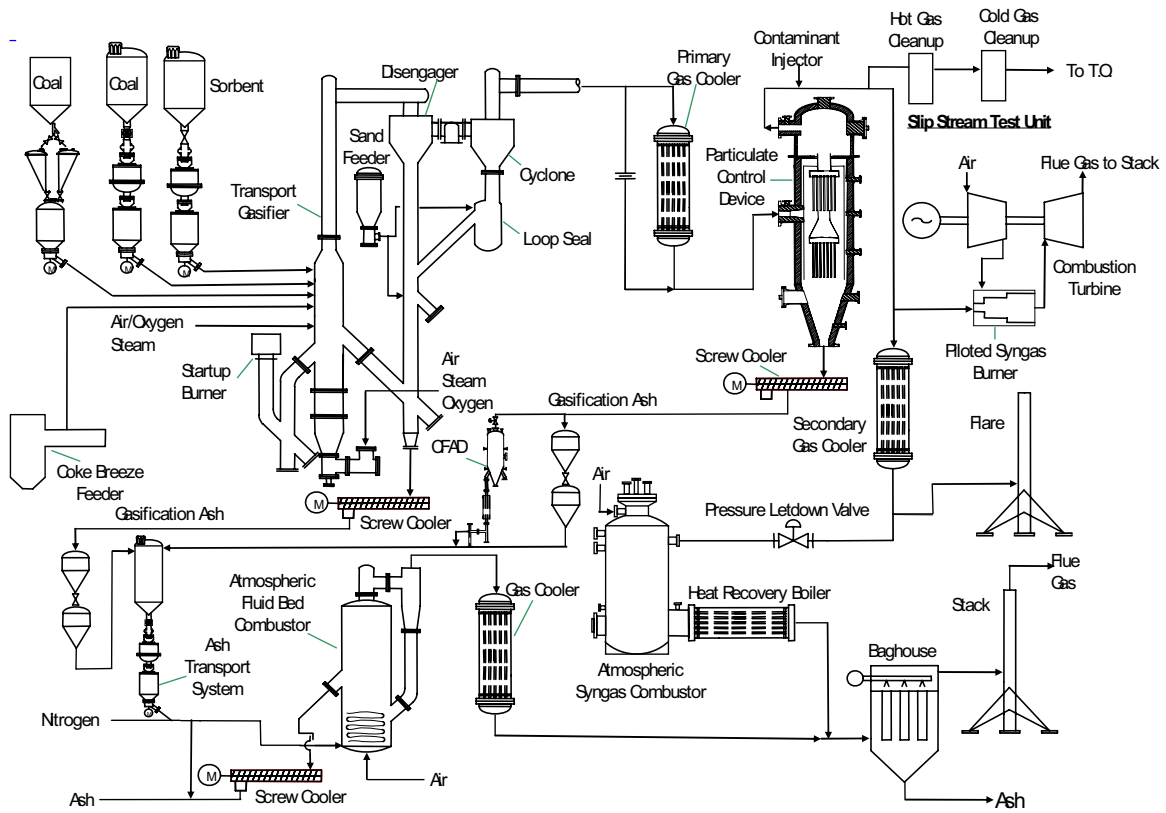


Figure 2.1-1 Flow Diagram of the Transport Gasifier Train

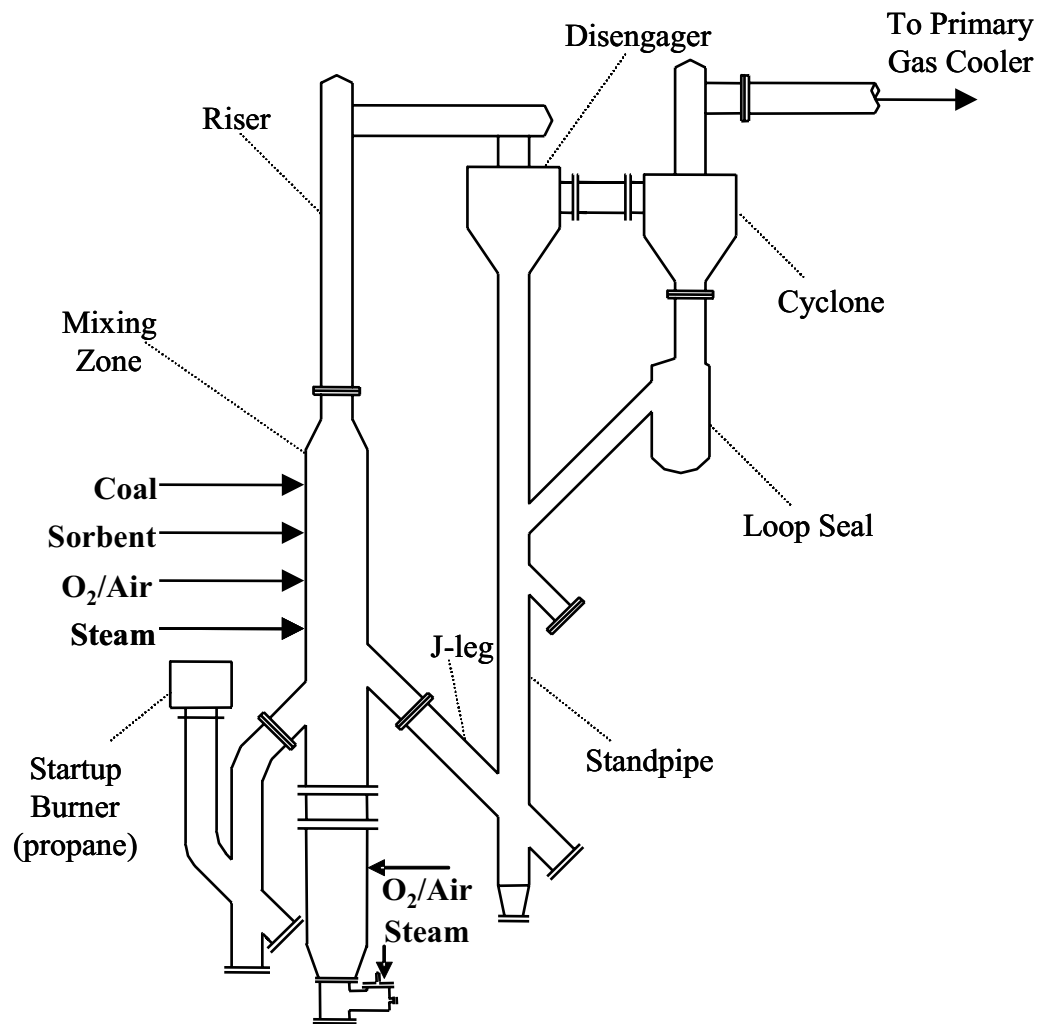


Figure 2.1-2 Transport Gasifier

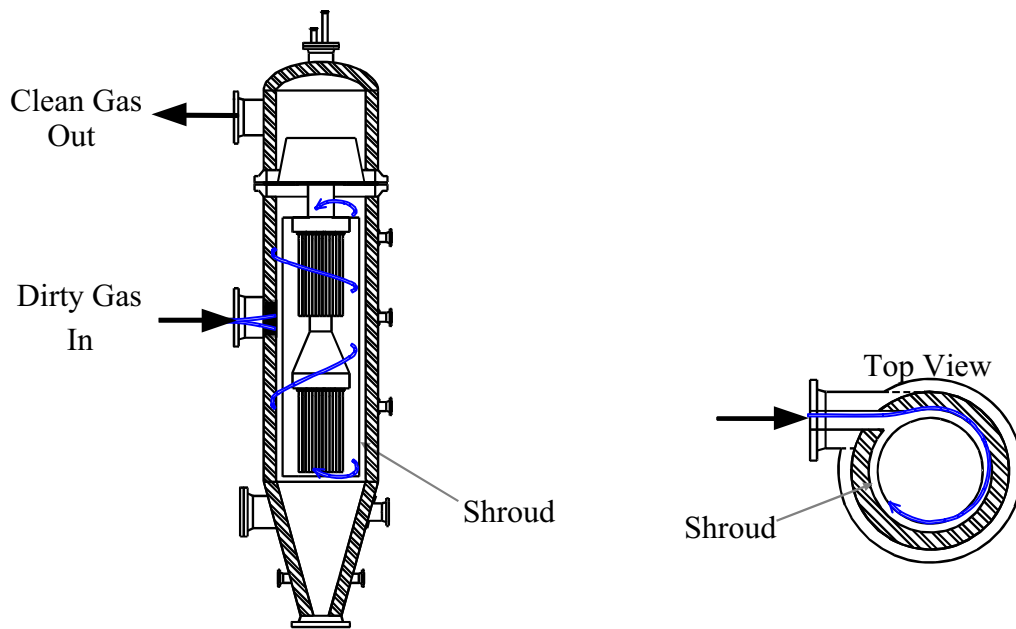


Figure 2.1-3 Siemens Westinghouse PCD

2.2 DETAILED TEST CAMPAIGN OBJECTIVES

TC15 was planned as a nominally 250-hour test run to evaluate gasifier and PCD operations using PRB subbituminous coal. The primary test campaign objectives were:

- Syngas Cleanup Commissioning – The syngas cleanup skid was commissioned and modifications were tested, including the new Fourier transform infrared spectroscopy analyzer (FTIR) and gas chromatograph (GC) analyzers with various sorbents and catalysts. The system demonstrated readiness for future fuel cell testing for TC16 by demonstrating the ability to remove sulfur.
- Oxygen Distribution System Commissioning – Attempts took place to commission two new oxygen nozzles in the mixing zone located just above the top air nozzles. The nozzles connected to the existing oxygen supply and distribution system and allowed for staging of the oxygen feed.
- Failsafe Device Testing – Online tests of the Ceramem ceramic failsafe and the Pall fuse were conducted, which included gasification ash injection into the clean side of two filters. The types of failsafe devices installed for gas exposure were the Pall fuse, CeraMem, PSDF-designed, and Westinghouse metal fiber failsafes.
- Higher Face Velocity Testing – High face velocity testing continued in TC15 with the removal of several filter elements. The PCD filtration performance and back-pulse cleaning effect at a higher face velocity were evaluated in TC15.
- Hot Gas Cleanup Commissioning – All parts of the gas cleanup train were commissioned. Testing investigated the integrity of various catalysts, including RSV-1, RSVLT-1, T-50 and G117, when exposed to syngas. The test plan demonstrated the operational reliability and measurement accuracy of the FTIR and GC with flame photometric detector (FPD).

Secondary objectives included the following:

- CFAD Testing – The development and testing of the CFAD system continued.
- PSB Testing – PSB and turbine system operations on syngas occurred.
- Wastewater Characterization – Samples of condensate collected from the syngas stream helped evaluate treatment needs for wastewater from Transport Gasifier-generated syngas.
- Evaluated the Effects of Gasifier Conditions on Process Performance – Efforts continued to evaluate the effects of different gasifier parameters such as the steam-to-coal ratio, CO-to-CO₂ ratio on carbon conversion, and product gas composition.
- Continued Sensor Development – Various tests evaluated thermowells and pressure differential measurements in the Transport Gasifier, including various materials of construction, detail design, and fabrication. One of the tests featured three new thermocouples at a 33° angle upward installation with various penetration depths. Efforts continued to evaluate nuclear density instruments for solids flow rates.
- Filter Element Testing – During TC15, the PCD tested only iron aluminide filters. Also, the seven Westinghouse inverted filter assemblies were tested further. Because the two Hastelloy-X filters (Grades P05 and P09) tested during TC14 showed

unacceptably high pressure drops across the filter media when flow tested, no testing of these particular Hastelloy filters occurred in TC15.

- Optimization of Back-Pulse Parameters – During the run, pressure drop measurements across an instrumented filter element were recorded for several back-pulse tank pressures and the actual pressure intensities across the filter media and dustcake were determined. In addition, back-pulse pressures were lowered for extended periods of time to assess the effect on baseline pressure drop and filter element cleaning. The purpose of this testing was to reduce the negative impact of back-pulsing on gasifier and PSB operations while maintaining adequate back-pulse intensity for filter cleaning.
- Particulate Characterization and Drag Evaluation – Gasification ash samples were collected in situ at the PCD inlet and thoroughly characterized to examine the effects of any changes in coal properties, carbon conversion, limestone addition, and operating conditions. The characterization included measurements of particulate loading, particle-size distribution, chemical composition, physical properties, and drag. As part of the continuing validation of the drag measurement procedures, the measurements made in the laboratory were compared to the transient drag values determined from the PCD pressure drop and particulate loadings.
- Analysis of Particulate Collection Performance – After establishing a clean baseline (less than 0.1 ppmw), a series of PCD outlet sampling runs were conducted in conjunction with the testing of the Ceramem failsafe and Pall fuse. These sampling runs helped to define the rate at which the Ceramem failsafe plugs and provided collection performance data on the Pall fuse to serve as a control.
- Ongoing Evaluation of PCME Particulate Monitor – Throughout TC15, the response of the PCME particulate monitor was evaluated to gain a better understanding of the instrument response and to detect any particle penetration through the PCD.
- Commission Minireactor Systems – The minireactor system was commissioned with simulated bottle gas for ammonia cracking using a selective catalyst.
- Evaluate the Effects of Operating Conditions on Cleanup Performance – Gas cleanup testing continued to evaluate the effects of operating parameters such as space velocity, bed height, and temperature on sorbent sulfur capturing capacity, exit sulfur concentration level, sulfur break through time, sorbent strength and elutriation, ammonia, and hydrocarbon cracking.

2.3 TEST CAMPAIGN SUMMARY

The activities that occurred during the outage preceding test run TC15 included 32 equipment modifications. The most significant of these are listed below:

- Two oxygen nozzles were added in the mixing zone to stage the oxygen in order to improve the gasifier temperature profile.
- Repairs to the gasifier cyclone removed the notch and replaced the damaged roof.

The following is a brief operating synopsis of TC15 from April 19 - April 29, 2004:

April 19, 2004

Once the cyclone refractory cure was complete, sand was added to the gasifier and the burner was relit at 12:45 p.m. The burner continued to heat the gasifier. Once the gasifier temperatures reached 1,100°F, coal-feed began. A controlled coal-feed rate of around 1,200 pph allowed the gasifier temperature to slowly climb to 1,700°F while in combustion mode. At this point, the coal-feed rate was increased to transition into gasification mode.

April 20, 2004

The gasifier continued to run at a steady coal-feed rate throughout the day. The CFAD system ran well at slow rates, removing about 250 pph of gasification ash from the PCD. Syngas flow to the gas cleanup system was started to allow inspection of the new FTIR analyzers. A back-pulse test was performed on the PCD in which the back-pulse pressure was varied from 150 to 400 psi above system pressure to assess the effect on baseline pressure drop and filter element cleaning. The testing of the developmental coal-feeder began delivering 3,500 to 5,000 pph of coal to the gasifier, while the original feeder fed a minimal amount of coal (around 500 pph). Later, coal fines packed in the lock hopper and the developmental feeder lost coal feed a few times. The developmental coal feeder was later shutdown and the coal-feed rate was increased on the original coal feeder. The main air compressor surged during the transition between the two feeders, causing a gasifier trip, but operations were quickly restored.

April 21, 2004

The developmental coal-feeder testing resumed. The CFAD ran at varying rates to remove the fines from the PCD. The developmental feeder was shutdown and the coal-feed rate on the original coal feeder was increased. The CFAD ran almost continuously throughout the night.

April 22, 2004

The PSB and CT test started in the morning. Syngas flow to the CT was initiated, resulting in a net heating value of about 90 Btu/scf. The circulation rate in the gasifier was increased to see its effect on the gasifier temperature profile. The cyclone exit temperature increased from 1,640 to 1,760°F while the two highest mixing zone temperatures decreased from 1,825 to 1,800°F. These changes occurred without any change in coal, air, or nitrogen flow rates. The PSB was shutdown after 10 hours of syngas flow. Work progressed on tuning and testing the fuzzy logic

portion of the single-loop gasifier temperature control. A number of coal-feeder step changes checked the response of the controller. The controller worked well except when the combination of changes affected the temperature profile in such a way that the controlled temperature measurement was no longer appropriate.

April 23, 2004

The Transport Gasifier pressure was reduced to 150 psig in preparation for oxygen-blown operation. The transition to oxygen-blown operations was smooth and took about an hour. The operating pressure was around 134 psig with the coal-feed rate around 3,000 pph. Solids circulation continued at the high rate, and the gasifier temperatures remained uniform. The syngas heating value increased to about 90 Btu/scf with a coal-feed rate of around 3,300 pph. The CFAD remained in operation. The pollution control and measurement (Europe) (PCME) online particulate monitor on the PCD outlet started showing an elevated reading. An outlet sample was collected that showed a concentration of 0.1 ppmw, which did not indicate a leak. The sample contained some solids and possible tar contamination. The condensate sample system continued to collect material during the day while the gasifier was in oxygen-blown mode. The new GC on the gas cleanup system was operational, and syngas flowed through the cleanup units to test the RVS1 sorbent. The sorbent testing was short-lived, however, due to the high moisture levels in the syngas flooding the GC columns. When testing the new oxygen nozzles in the mixing zone, a temperature excursion occurred, forming an agglomeration in the UMZ. The agglomeration prevented oxygen flow to the two new nozzles in the UMZ. The restriction also prevented any further testing on the developmental feeder. Testing on the CFAD system and condensate collecting continued.

April 24, 2004

The gasifier pressure was increased to 145 psig while the temperatures and riser velocity remained roughly the same. Steady state conditions were maintained to allow testing of the CeraMem failsafe with the online failsafe tester. Two PCD outlet particulate samples were taken to assess the collection performance of the CeraMem failsafe. During collection of the first sample, the sampling line plugged and sampling was stopped prematurely. The outlet loading from the first sample was roughly estimated at approximately 0.5 ppmw. The second sample was collected about 5 hours later and the outlet loading was approximately 0.17 ppmw. The CFAD system ran well, achieving various solids flow rates. Hot gas desulfurization testing also took place at this time. The slipstream unit was able to drop the H₂S content to around 20 ppm. Oxygen flow to the two new nozzles in the UMZ was still unachievable due to the restriction that had occurred earlier.

April 25, 2004

CFAD testing continued to determine the various discharge rates of the fines removed from the PCD. The gasifier ran at a slightly higher pressure of 150 psig in order to accumulate cyclone efficiency data at a lower velocity. The coal-feed rate remained around 4,000 pph, while the syngas lower heating value was close to 90 Btu/scf. The new UMZ nozzles cleared, allowing oxygen to flow to the UMZ. The temperature profile remained unchanged, however. PCD operations were smooth throughout the day and the pressure drop was stable with a lower back-pulse pressure of 150 psi above system pressure. The CFAD continued to run well with

discharge rates as high as 1,700 pph. The increase in gasifier pressure from 144 to 150 psig resulted in solids accumulating in the standpipe. Thus, the lower velocity appeared to improve the cyclone efficiency. Due to gasifier problems with solids circulation caused by a standpipe bubble, the gasifier was returned to air-blown operation to avoid excessive temperatures during the unstable condition.

April 26, 2004

Once gasifier operations steadied, the transition back to oxygen-blown operations took place. During the transition, some of the temperatures in the mixing zone were higher than those seen in past runs. Some temperature spikes were seen on the top plenum of the PCD, but not enough to cause a rate-of-change alarm. The gasifier lost circulation once again requiring the cessation of coal-feed and the relighting of the start-up burner. Restoring circulation was difficult. The removal of the entire standpipe solids inventory—along with adjustments in J-leg flows—proved essential in removing a portion of the restriction, allowing some solids circulation to take place. The high aeration flows unplugged the J-leg, and the test run was able to continue. The material removed from the standpipe contained several small agglomerations. Once the standpipe was mostly cleaned, sand was added to increase gasifier inventory and to reestablish solids circulation. After establishing solids circulation, the start-up burner heated the gasifier to around 1,100°F, and coal-feed resumed. Three hours after restarting the coal-feeder, the transition to oxygen-blown operation occurred. The transition was smooth, but the circulation still appeared to be substandard. The gasifier experienced a few high-temperature trips and one extended period of poor circulation, but was able to continue operating in oxygen-blown mode.

April 27, 2004

The gasifier continued to experience problems with circulation, as indicated by temperature readings. For most of the day, the mixing zone temperature was around 1,850°F, while the cyclone outlet temperature was only around 1,450°F. Lowering the coal-feed rate helped the gasifier achieve a slightly more uniform temperature profile. As the coal grind size decreased, the coal-feeder speed was increased, but the temperature profile remained slightly abnormal.

April 28, 2004

Circulation problems continued causing the temperature profile to become less stable. Reverse flow occurred and conditions required the transition back to air-blown operation. Eventually, the cessation of coal-feed was necessary to allow for more aggressive attempts to restart circulation by dramatically adjusting flows. Later, circulation in the gasifier was restored, but at a limited rate. The initial attempts to light the start-up burner failed; therefore, the unit was taken offline to inspect the burner. The flame rod and flame scanner were in satisfactory condition, but the igniter required replacement. Once the unit was back at 60 psig, the start-up burner lighting sequence progressed. The heating of the gasifier continued throughout the day. Once the gasifier was at sufficient temperature, coal feeding was resumed. The gasifier operated in air-blown mode at a low coal-feed rate and low pressures.

April 29, 2004

Due to the uneven temperature profile in the gasifier caused by lack of solids circulation, coal-feed was stopped and the start-up burner was lit again. After gasifier temperatures were increased, coal feed was resumed. However, the solids circulation remained low, resulting in an uneven temperature distribution. Since the gasifier operations were unsteady, the unit was taken offline and the test run ended.

Typical operating conditions for the Transport Gasifier and the PCD can be found in [Table 2.3-1](#).

Table 2.3-1

TC15 Typical Operating Conditions for the Transport Gasifier and the Particulate Control Device

Transport Gasifier	
Startup Bed Material	Sand, ~ 120 μm
Startup Fuel	Coal
Fuel Type	Powder River Basin
Fuel Particle Size (mmd), μm	220-350
Average Fuel Feed Rate, pph	3,000-5,000
Maximum Gasifier Temperature, $^{\circ}\text{F}$	1800-1850
Mixing Zone Pressure, psig	110-150 (O_2), 200-230 (air)
Riser Gas Velocity, fps	35 - 55
Standpipe Level, inH ₂ O	45-175
Total Gas Flow Rate, pph	14,000 - 22,000
Oxygen/Coal Mass Ratio, lb/lb	0.6 - 0.8
Oxygen/Steam Mass Ratio, lb/lb	0.6-2.0
Steam/Coal Mass Ratio, lb/lb	0.3 to 1.1
Particulate Control Device	
PCD Temperature, $^{\circ}\text{F}$	700-950
PCD Inlet Loading, ppmw	9,700-29,700
PCD Outlet Loading, ppmw	< 0.1
PCD Baseline Pressure Drop, inH ₂ O	60 to 120
Filter Element Type (number)	Pall FEAL (73)
Filtration Area, ft^2	207.3
Face Velocity, ft/min	3.0-5.0
Pulse Valve Open Time, sec	0.2
Pulse Time Trigger, min	5-15
Pulse Pressure, Top Plenum	150-320 psi above System Pressure
Pulse Pressure, Bottom Plenum	150-320 psi above System Pressure

2.4 DETAILED INSPECTIONS/CONCLUSIONS

2.4.1 Transport Gasifier

The post TC15 inspection showed the gasifier to be in generally good shape with two major exceptions. The first was a significant deposit in the transition from the mixing zone to the riser. The second was a smaller deposit in the gasifier J-leg.

The riser and mixing zone were inspected with the boroscope. The transition region from the mixing zone to the riser had a significant deposit. From above the original coal-feed nozzle to just below the lowest level of the combustion, secondary air nozzles were almost completely blocked. The restriction also covered the developmental coal-feeder injection point. Below the coal-feed nozzle, there were deposits on the wall but with a large opening to the center.

In the riser, there were scattered deposits above the main deposit up to the second level of secondary air nozzles (approximately 30 feet above the coal injection point). Above this, the riser was clear. There were also deposits around the second level of secondary air nozzles. On top of these thermowells was a layer of deposits. The deposits were rather small and round and were in a loose pile on top of the thermowells. The thermowells could not be seen but there was space enough to pass the boroscope.

Below the deposits in the UMZ, the mixing zone (including the LMZ) was rather clean.

The riser crossover had a small, loose pile of material at the south (riser) end. This was not a deposit and is a common feature during inspections. The disengager and the cyclone looked to be in good shape.

The standpipe did not have any deposits, and the cracks seen were consistent with previous test runs. The loop seal was clean except for the normal, soft deposits on the walls of the downcomer. These deposits were about where the top of the solids level should be during operation. The J-leg had a small deposit at the very bottom, just below the lowest set of aeration nozzles. It was fairly soft and blocked only a small part of the flow path.

The burner and burner J-leg were inspected with the boroscope. They were found to generally be in good shape. Some loose deposits, likely from the mixing zone, were found in the bottom of the J-leg.

The primary gas cooler was in rather good shape. There were a few plugged tubes and a thin layer of dust on the tubesheet. The secondary gas cooler had a collection of heavy organic compounds in the cone but was otherwise clean.

2.4.2 Particulate Control Device

The PCD was inspected after shutdown and there was no evidence of char bridging or filter element failures. A clean shutdown, in which the PCD was back-pulsed for 24 hours after coal-feed was terminated, left no transient dustcake and a very thin residual dustcake. The dustcake on the top plenum was generally thicker than seen in previous test runs. Although the dustcake

had a strong odor, it was easily brushed off and the filter element surface appeared dry and smooth. However, the rod on the PCME particulate monitor was covered by black tar.

The seven Siemens Westinghouse inverted filter assemblies were removed from the top plenum and flow tested. There was no significant amount of material found inside any of the filters. Flow test results for these filter assemblies revealed that the pressure drop at a face velocity of 3 ft/min ranged from 6.8 to 7.8 inches of water. This low pressure drop indicated that the filter assemblies performed well during the test run. The failsafes above these elements were also removed and flow tested. The failsafes were in good condition and were not plugged.

3.0 PERFORMANCE

3.1 TRANSPORT GASIFIER PERFORMANCE

3.1.1 Overview

- The coal moisture ranged from 18.4 to 24.2 percent. The coal ash composition remained relatively constant during the test run.
- The standpipe solids had a CaS content of around 0.1 percent and a carbon content of less than 3 percent. The standpipe solids particle size did not reach steady state values due to frequent sand addition, and the standpipe solids bulk density oscillated between 79 and 94 lb/ft³.
- The total calcium, CaS, and SiO₂ concentrations of the PCD inlet in situ samples were consistent with FD0520/FD0540 sample concentrations.
- The PCD fines and standpipe solids sulfur contents indicate minimal Transport Gasifier sulfur capture.
- During air-blown gasification, the raw LHVs at the exit of the gasifier were between 26 and 74 Btu/scf, resulting in projected heating values at the turbine inlet of a commercial gasifier of between 106 and 135 Btu/scf. The heating values were slightly lower in TC15 than in previous test runs due to a lower coal-feed rate. Lower circulation rates limited the coal-feed rate to lower than usual, reducing the heating value.
- In oxygen-blown gasification, the raw LHVs at the exit of the gasifier ranged from 80 to 98 Btu/scf, resulting in projected heating values of 256 to 275 Btu/scf.
- The syngas molecular weight was between 26.5 and 27.6 pounds/pound-mole in air-blown and between 23.7 and 24.4 pounds/pounds-mole in oxygen-blown operation.
- Total reduced sulfur (TRS) concentrations, mostly H₂S, were between 268 and 410 ppm for air-blown operation and between 574 and 709 ppm for oxygen-blown operation. Coal sulfur content was 0.24-percent sulfur. No sorbent was added.
- The carbon conversion was between 91 and 98 percent for both air- and oxygen-blown gasification.
- The raw cold gas efficiency ranged from 38 to 61 percent during air-blown operations and from 61 to 64 percent during oxygen-blown operations. The commercially projected efficiency was between 65 and 77 percent for the air-blown periods and from 83 to 91 percent during oxygen-blown operations.
- The hot gas efficiency ranged from 70 to 87 percent for the air- and oxygen-blown periods.

The test run had 17 periods of steady operation between April 20 and April 29, 2004. These periods are given in [Table 3.1-1](#). The operating periods had a cumulative time of about 47 hours, which was about 23 percent of the total TC15 on-coal operation time. The first seven periods and the last period occurred during air-blown operation. Periods TC15-7 through TC15-15 were oxygen-blown. There was no sorbent used during TC15.

[Table 3.1-2](#) lists the TC15 operating conditions, including the coal-feed rate, the riser exit temperature, the system pressure, the PCD inlet temperature, the PCD solids collection rate, the air rate, oxygen rate, syngas rate, steam rate, and nitrogen rate of the steady operating periods.

The system pressure ranged from 127 to 214 psig for the air-blown periods and from 111 to 150 psig during oxygen-blown operations. Riser exit temperatures were between 1,409 and 1,753°F. The exit temperatures were lower than in previous test runs due to lower circulation rates. Steam flow rates were between 500 and 1,000 pph in air-blown mode and between 1,900 and 3,400 pph in oxygen-blown mode.

3.1.2 Gas Composition

During the test run, Transport Gasifier and syngas combustor outlet gas analyzers were continuously monitored and recorded by the Plant Information (PI) system. Six in situ samples of syngas were taken during PCD outlet loading sampling and measured for moisture content. Plotted on [Figure 3.1-1](#) is the syngas moisture analyzer (AI475H), the in situ H₂O concentrations versus time, and moisture content based on a correlation of temperature and the in situ data. The moisture content during air-blown gasification was around 10 percent. During oxygen-blown gasification, the moisture content increased to over 20 percent due to higher steam-flow rates and lower syngas rates.

The H₂O concentrations calculated for the operating periods (based on the correlation) are given in [Table 3.1-3](#). They ranged from 9.1 percent to 10.7 percent in air-blown mode and were 21.6 percent to 28.5 percent for the oxygen-blown periods. Based on these moisture concentrations, the estimated wet syngas compositions for the TC15 operating periods are given in [Table 3.1-3](#) and shown on [Figure 3.1-2](#). Also shown in [Table 3.1-3](#) are the syngas molecular weights for each operating period.

The CO concentration typically ranged from 6.3 to 11.4 percent during air-blown gasification, except for TC15-17 when the CO content was only 2.3 percent due to a low coal-feed rate. The CO content was between 6.7 and 11.5 percent during the oxygen-blown periods. CO concentrations were generally higher in oxygen-blown mode due to decreased nitrogen dilution.

The H₂ concentration fluctuated between 4.8 and 7.3 percent during the initial air-blown periods. After the transition to oxygen-blown gasification, the hydrogen concentration increased to between 11.8 and 14.0 percent. The increase was due to the decrease in nitrogen dilution and the higher steam flow rates that shifted equilibrium conditions in the gasifier to produce extra hydrogen. During the last air-blown period, TC15-17, the hydrogen content was only 3.1 percent because of the low coal-feed rate during this period.

The CO₂ concentration ranged from 8.4 to 9.4 percent during air-blown gasification. During the two oxygen-blown periods, the CO₂ concentration increased to between 13.0 and 14.7 percent due to the decrease in nitrogen dilution and the water-gas shift reaction.

During air-blown operations, the CH₄ concentration was between 0.6 and 1.9 percent. The methane content ranged from 1.9 and 2.8 percent during oxygen-blown mode.

The C₂⁺ concentration was negligible for all of the test periods, with the exception of TC15-16, when it was 0.1 percent.

During TC15, one of the gas analyzers detected argon present in the syngas. It ranged from 0.0 to 0.5 mole percent.

The wet syngas molecular weight and nitrogen concentration are plotted on [Figure 3.1-3](#). The air-blown gasification molecular weights ranged between 26.5 and 27.6 pounds per pound mole. The oxygen-blown molecular weights were between 23.7 and 24.4 pounds per pound mole. The decrease in molecular weights during the oxygen-blown period was due to higher concentrations of lower molecular weight compounds such as H_2 and H_2O .

The CO/CO_2 ratios were calculated from the gas data for each operating period and are listed in [Table 3.1-3](#). The CO/CO_2 ratio varied from 0.26 to 1.35, with the lower ratios occurring during periods of low coal-feed rates.

The main sulfur species in coal gasification are H_2S and carbon oxysulfide (COS) with other sulfur compounds, such as CS_2 , present in small quantities. The TRS is the sum of the compositions of all sulfur species in the syngas. When combusted in the syngas combustor, all sulfur compounds present in the syngas are converted to SO_2 . The wet H_2S concentration (measured by AI419J) and the SO_2 content at the syngas combustor exit (measured by AI476N) are plotted on [Figure 3.1-4](#). Also plotted on the graph is the calculated syngas TRS concentration derived from the SO_2 content at the syngas combustor exit. The wet H_2S concentration and the syngas TRS concentration are also listed in [Table 3.1-4](#). The AI419 analyzers measure the gas composition on a dry basis, so the values from AI419J were corrected to include moisture. The syngas combustor SO_2 analyzer, AI476N, measures the total sulfur emissions exiting the system.

The TRS concentration was 304 ppm at the beginning of the test run, and it ranged between 268 and 410 ppm during the air-blown test periods. During oxygen-blown mode, the TRS increased to between 478 and 661 ppm due to less nitrogen dilution. No limestone feed occurred during the test run; therefore, the only calcium present in the gasifier to capture sulfur came from the PRB coal ash. In previous PRB test runs, the use of sorbent has had little effect on syngas sulfur concentrations.

The TRS concentrations are plotted against the wet AI419J data on [Figure 3.1-5](#). All of the data points for air- and oxygen-blown operations are outside 30 percent of the TRS values. Note that operating in oxygen-blown mode also increases the TRS emissions due to less TRS dilution by the nitrogen in the air. TRS concentrations should be higher than the H_2S concentrations due to COS and CS_2 present in the syngas, but since the disparity between the H_2S analyzer data and the TRS data was over 30 percent (and the H_2S analyzer occasionally read higher than the calculated TRS), the discrepancy is likely due to the analyzer reading improperly at low H_2S concentrations.

The reduced nitrogen present in the syngas is mostly in the form of ammonia. Hydrogen cyanide (HCN) is also present. When these materials are combusted in the syngas combustor, they produce NO_x . [Table 3.1-4](#) shows the ammonia concentrations in the syngas. The ammonia concentration ranged from 752 to 1,985 ppm in air-blown mode and from 1,779 and 3,988 ppm in oxygen-blown mode.

3.1.3 Syngas Heating Values

Raw Syngas Heating Values

The raw syngas lower heating value (LHV) for each operating period was calculated and is listed in [Table 3.1-3](#) and plotted on [Figure 3.1-6](#). All raw LHVs are on a wet basis.

The LHV was calculated using the formula:

$$\text{LHV(Btu/SCF)} = \frac{\left\{ 275 \times (\text{H}_2\%) + 322 \times (\text{CO}\%) + 913 \times (\text{CH}_4\%) + 1641 \times (\text{C}_2+\%) \right\}}{100} \quad (1)$$

During air-blown gasification the LHV typically ranged from 57 to 74 Btu/scf. Two exceptions occurred during steady state periods, TC15-7 and TC15-17, when the heating values were only 39 and 26 Btu/scf. During these periods, the coal-feed rates were low, and the nitrogen-dilution rates per Btu of coal fed were high. The LHV ranged from 80 to 98 Btu/scf during the oxygen-blown periods.

Past test runs have indicated that the most significant impact on LHV are the coal- and steam-feed rates (see Figure 4.5-5 of the TC06 Final Report). As the coal rate increases, the syngas production rate increases while the aeration and instrument purge nitrogen flow rates remain constant. Therefore, the nitrogen constituent of the syngas decreases (less dilution), and the syngas LHV increases. During oxygen-blown gasification, very little nitrogen enters the gasifier with the oxidant, increasing the LHV; however, the higher steam flows in oxygen-blown mode negate this effect.

Increasing the steam flow decreases the LHV by diluting the syngas with moisture. One way to combine the effects of changing in the mode of operation, as well as, the change in steam- and coal-feed rates, is to determine the overall percent of oxygen of all the gas that enters the Transport Gasifier. The overall percent O₂ is calculated by the following formula:

$$\text{Overall Percent O}_2 = \frac{.21 * \text{air} + \text{oxygen}}{\text{air} + \text{oxygen} + (\text{pure nitrogen}) + \text{steam}} \quad (2)$$

All flows are in moles per hour. At the PSDF, a large amount of pure nitrogen is fed to the gasifier for instrument purges, coal and sand transport, and equipment purges. During air-blown gasification, about half of the nitrogen flowing through the gasifier comes from these pure nitrogen flows and the remainder comes from the nitrogen in the air. When operating in oxygen-blown gasification, the nitrogen in the gasifier is predominantly due to the pure nitrogen flows.

The TC15 percent O₂ overall values are listed in [Table 3.1-3](#). The values range from 10.1 to 14.0 percent O₂ in air-blown mode and from 15.1 to 18.9 percent O₂ in oxygen-blown mode. The overall percent O₂ is higher in oxygen-blown gasification due to less nitrogen dilution.

The TC15 raw LHV data is plotted against overall percent O₂ on [Figure 3.1-7](#). As the overall percent O₂ increases, the LHV also increases. For comparison, the curves of previous PRB data (TC06, TC07, TC08, TC10, TC12, TC13, and TC14), previous Hiawatha bituminous data (TC09), Falkirk lignite (TC11), and Freedom lignite (TC13) are included. The general trend for TC15 compares well with PRB data generated over the past several runs. The two data points at lower coal-feed rates yielded LHVs much lower than the general trend. The heating values were higher for PRB than for either of the lignites at the same overall percent O₂, but lower than the Hiawatha bituminous.

Projected Syngas Heating Values

A commercial-sized Transport Gasifier will produce syngas of a higher quality than the PSDF gasifier due to:

- The use of recycled gas rather than nitrogen for aeration and PCD back-pulsing.
- A lower heat loss per pound coal gasified because of a lower surface area to volume ratio.
- A smaller number of instruments and instrument purges.
- A cold gas clean-up train to remove contaminants and moisture from the syngas.

For details on the projected LHV calculation, see [Appendix 5](#).

The projected LHV and adjusted syngas composition data for each operating period are given in [Table 3.1-5](#) at the gas turbine inlet. The projected LHV was between 106 and 135 Btu/scf for air-blown operation and was between 256 and 275 Btu/scf during the oxygen-blown periods.

For comparing the raw LHVs with the projected LHVs, an equivalent to the projected overall percent O₂ is defined as:

$$\text{Projected Overall Percent O}_2 = \frac{.21 * (\text{projected air}) + (\text{projected oxygen})}{(\text{projected air}) + (\text{projected oxygen}) + (\text{projected steam})} \quad (3)$$

All flow rates are expressed as moles per hour.

The projected LHV is plotted against the projected overall percent O₂ on [Figure 3.1-8](#). The linear fits of the projected PRB LHV data, the projected Falkirk lignite LHV data, the projected Hiawatha bituminous LHV data, and the projected Freedom lignite data from previous test campaigns are also shown on [Figure 3.1-8](#). In oxygen-blown mode, the TC15 projected LHV data was slightly higher than the average projected PRB oxygen-blown data from previous test campaigns. In air-blown mode, the TC15 values were consistent with those of previous PRB tests.

3.1.4 Gasifier Solids Analyses

During TC15, the solid samples were taken from the following locations:

- The coal-feed system (FD0210).

- The Transport Gasifier standpipe (coarse gasification ash).
- The Transport Gasifier loop seal downcomer.
- The PCD fine solids lock-hopper system (FD0520).
- The new CFAD or FD0540.

In situ solids samples were also collected from the PCD inlet. All solid samples were analyzed for chemical composition and particle size.

PRB coal was the fuel for the entire test run. [Table 3.1-6](#) gives the average proximate, ultimate, and ash mineral analyses of the PRB coal, as well as, the average molar ratios for coal calcium-to-sulfur ratio (Ca/S) and the standard deviation for the samples analyzed as sampled from FD0210. The fuel carbon and moisture contents are shown in [Figure 3.1-9](#). Both the moisture and the carbon content of the PRB coal remained essentially constant during TC15.

[Figure 3.1-10](#) shows the coal sulfur and ash as sampled from FD0210 during TC15. The sulfur level remained between 0.22- and 0.27-weight percent. The ash content was typical for PRB coal and remained mostly constant, between 4.8- and 5.5-weight percent.

The higher heating value (HHV) and the LHV of the coal is given in [Figure 3.1-11](#) with the TC15 average value given in [Table 3.1-6](#). The coal HHV is determined using a bomb calorimeter. The calorimeter condenses all the coal combustion moisture as liquid water. The LHV is calculated by subtracting the heat of vaporization of the coal moisture from the HHV. Since heat recovery steam generators do not recover the coal syngas moisture heat of vaporization, the LHV is a more useful measure of the coal heating value. The LHVs for the PRB coal ranged from 8,400 to 9,000 Btu/scf during the test run.

The chemical compositions of the solid compounds produced by the Transport Gasifier were determined based on the chemical analysis and the following assumptions:

1. All carbon dioxide measured is from CaCO_3 , hence moles CO_2 measured = moles CaCO_3 .
2. All sulfide sulfur measured is from CaS .
3. All calcium not taken by CaS and CaCO_3 is from CaO .
4. All magnesium is from MgO .
5. Total carbon is measured, which is the sum of organic and inorganic (CO_2) carbon. The organic carbon is the total carbon minus the inorganic carbon (CO_2).
6. Inerts are the sum of the Fe_2O_3 , BaO , P_2O_5 , Na_2O , K_2O , and TiO_2 concentrations.

Both elemental sulfur (ultimate analysis) and ash inerts sulfur contents were measured.

[Table 3.1-7](#) gives the TC15 standpipe solids analyses. These solids recirculate through the mixing zone, riser, and standpipe. Typically, the properties of these solids change slowly with time. Due to low solids capture efficiency, the coarse standpipe spent solids transporter system (FD0510) operated infrequently. The solids samples normally come directly from the standpipe via the standpipe sampler; however, the standpipe sampler was out of service during the majority of TC15, so the samples came from the FD0510 feed system. Since the FD0510 feeder is

directly below the standpipe, the samples should be fairly representative of the material in the standpipe.

During the test run, the gasifier lost solids circulation a few times. To maintain solids levels, the FD0220 feeder was used several times to add sand. Due to the frequent additions of sand, the standpipe solids' silica content was abnormally high, typically between 79- and 91-weight percent. [Figure 3.1-12](#) shows the standpipe SiO_2 , CaO , and Al_2O_3 contents versus run time, showing the high silica concentration. Based on previous test campaigns, the standpipe solids did not reach a steady state value in TC15 (~50 percent SiO_2).

The standpipe solids organic carbon content is plotted on [Figure 3.1-13](#). The organic carbon is the total carbon in the solids minus inorganic carbon measured as CO_2 . Based on previous experience, the standpipe organic carbon content is a very inaccurate measurement because the value comes from a difference of two small values that are nearly equal. The standpipe organic carbon content was between 0.42- and 2.21-weight percent. In general, the standpipe solids' carbon content was higher in TC15 than it was in previous test campaigns due to low circulation rates. The low circulation rates caused an increase in temperature drop through the gasifier which in turn resulted in a larger amount of unburned carbon circulating in the gasifier loop.

The standpipe solids CaS content was negligible for all of the standpipe samples. The calcium in the PRB coal ash captured a minimal amount of sulfur which is consistent with previous PRB test data. The standpipe CaCO_3 was between 1.4 and 3.4 percent for TC15. About 66 percent of the calcium in the standpipe solids was calcined to CaO .

[Table 3.1-8](#) lists the solids sample analysis for the loop seal. The solids from the loop seal are the solids that pass through the disengager with the syngas, but are captured in the cyclone. After the cyclone captures the loop seal solids, they flow back to the standpipe where they join the solids falling from the disengager.

[Figure 3.1-14](#) shows the CaO , SiO_2 , and Al_2O_3 contents of the loop seal samples. The loop seal solids were lower in SiO_2 content than were the standpipe solids. The loop seal SiO_2 content began the test run at 86.3 percent and declined to 40.6 percent as carbon replaced the silica. CaO and Al_2O_3 contents remained less than 10 percent the entire test run.

[Figure 3.1-15](#) shows the organic carbon (total carbon minus CO_2 carbon) and CaCO_3 contents for the loop seal solids. The carbon content of the loop seal solids was considerably higher in the loop seal than in the standpipe (1 to 52 percent versus 0 to 2 percent), because a greater percentage of the smaller carbon particles were collected by the cyclone rather than by the disengager. The loop seal CaCO_3 slowly increased from 1.8 to 4.6 percent. The MgO , Fe_2O_3 , and other inert contents are not plotted, but they follow the same trend as the Al_2O_3 due to the frequent sand additions. The loop seal solids CaS content was negligible for all samples.

The complete solids analysis, as well as, organic carbon content for the PCD fines samples, is given in [Table 3.1-9](#). In situ PCD inlet particulate solid samples were also analyzed.

[Figure 3.1-16](#) plots the organic carbon for the PCD solids sampled from the spent fines feeder (FD0520) and the CFAD system (FD0540), as well as, for the in situ samples. Since FD0540 was in use for the majority of the run, all but two of the samples (indicated on [Figure 3.1-16](#) and all other PCD fines plots) came from FD0540. The first PCD sample was taken about 15 hours

after coal feed began and the organic carbon content was only 9.7 percent. The carbon content remained between 17.5 and 50.5 percent for the rest of the test run.

Figure 3.1-17 and Table 3.1-9 also show the amounts of SiO_2 and Al_2O_3 in the PCD solids as sampled from FD0520 and FD0540. The in situ solids concentrations for SiO_2 and Al_2O_3 are also plotted in Figure 3.1-17. The SiO_2 PCD fines' concentrations are a function of the efficiency of the disengager and cyclone, as well as, the SiO_2 concentration of circulating solids in the gasifier. The SiO_2 in the PCD fines is made of fresh coal ash SiO_2 , startup sand SiO_2 that is being lost from the gasifier and additional sand that is added to the gasifier. The SiO_2 PCD fines concentration widely fluctuated between 20 and 59 percent due to frequent sand additions. Typically, the SiO_2 content is 30 to 40 percent at steady state.

Since only a minimal amount of Al_2O_3 is in the startup sand, the PCD fines Al_2O_3 content comes predominantly from the coal ash. The PCD fines Al_2O_3 concentration remained low, between 5 and 11 percent, for the majority of the test run.

Figure 3.1-18 shows the calcium carbonate and calcium sulfide concentrations in the PCD solids as sampled from FD0520 and FD0540. The concentrations for CaO , CaS , and CaCO_3 are also listed in Table 3.1-9. Also plotted on Figure 3.1-18 are the calcium carbonate and calcium sulfide concentrations for the in situ solids samples. All of the in situ sample CaS concentrations agreed well with FD0520 and FD0540 solids calcium sulfide concentrations. Some of the in situ sample calcium carbonate contents agreed with the FD0520 and FD0540 samples. The PCD fines calcium carbonate concentration fluctuated from less than 3.4 to 7.4 percent. Since no limestone feed occurred during the test run, all of the PCD fines calcium came from the PRB coal ash. The PCD fines CaS concentration for PRB was 1.2 percent or less for the entire test run, indicating very low sulfur capture.

The PCD fines calcination is defined as:

$$\% \text{ Calcination} = \frac{M\% \text{ CaO}}{M\% \text{ CaO} + M\% \text{ CaCO}_3} \quad (1)$$

The PCD fines calcination data are plotted on Figure 3.1-19. The PCD fines calcination ranged from 49 to 82 percent, as was typical in previous PRB runs.

The calcium sulfation is defined as:

$$\% \text{ Sulfation} = \frac{M\% \text{ CaS}}{M\% \text{ CaO} + M\% \text{ CaCO}_3 + M\% \text{ CaS}} \quad (2)$$

The PCD fines sulfation is plotted on Figure 3.1-19 with the PCD fines calcination. The PCD fines sulfation varied between 1 and 8 percent for most of the test run. These values indicate that only a minimal amount of sulfur capture took place.

Solids Sample Comparison

The following comparison of the analysis of the standpipe solids, the loop seal downcomer solids, and the PCD fines solids shows how the solids compositions change throughout the process.

Figure 3.1-20 compares the organic carbon content of the standpipe, loop seal, and PCD fines solids samples. The PCD solids carbon content ranged from 10 to 51 percent, with a high variability due to sand addition. The loop seal organic carbon content ranged between 1 and 52 percent, and the standpipe solids carbon content was between 0 and 2 percent. This data seems to indicate that the carbon is contained in small particles which are only partially captured by the disengager. The cyclone separates a larger portion of the carbon, and the PCD collects the rest. In general, the loop seal organic carbon data fell between the standpipe and PCD fines data, with the exception of the final loop seal data point.

Figure 3.1-21 compares the calcium concentration between the standpipe, the loop seal, and the PCD fines solids samples. The calcium content was always highest for the PCD solids, ranging from 3 to 9 percent. The loop seal solids calcium content was usually higher than that of the standpipe solids, with the exception of the final data sample that was taken during a period of unstable solids circulation. The standpipe calcium content remained relatively constant, while the PCD and loop seal calcium content changed, probably due to the dilution caused by the frequent sand addition and unstable solids circulation. The higher PCD fines calcium content indicated that the calcium was concentrated in the smaller particles that passed through the disengager and cyclone. (Note that the calcium is distributed between the compounds CaO , CaCO_3 , and CaS .)

The silica entering the process primarily remains in the gasifier, since the sand particle size is greater than that of the standpipe solids. Figure 3.1-22 shows that the standpipe solids had the highest silica content, followed by the loop seal. The PCD solids had the lowest silica content, but it fluctuated over a larger range, becoming higher during periods of poor solids collection performance and sand addition. As the gasifier operates, sand eventually breaks down by attrition and exits through the cyclone before being collected by the PCD. Typically, the sand loss through the cyclone is minimal, and the replacement of sand with bed material is a slow process. In TC15, sand feed was more frequent than in previous test runs, and the test run was short, so the silica inventory in the gasification loop never reached steady state.

Solids Particle Size

The TC15 Sauter Mean Diameter (SMD) and Mass Mean Diameter (MMD) particle sizes of the coal sampled from FD0210 are plotted on Figure 3.1-23. The PRB coal SMD particle size was fairly steady, from just under 159 μm to 308 μm . The average SMD for the PRB was 216 μm , with a standard deviation of 40 μm .

Figure 3.1-24 plots both the coal-feed percent above 1,180 μm (coarse particles) and percent below 45 μm (fines). A large amount of particles above 1,180 μm increases the difference between the SMD and the MMD, because the SMD is a surface area average. Therefore, the larger particles with less surface area per pound have a weaker effect on the SMD than the MMD,

where the larger particles skew the MMD due to their higher weight per particle. The average percent above 1,180 μm for TC15 was 7.5 percent with a standard deviation of 2.1 percent. The percent above 1,180 varied during the entire test run, between 4 and 11 percent. The high spikes in coarse solids correspond to the high MMD particle size.

In past testing, a high fines content in the feed coal resulted in an increased number of coal-feeder outages due to the packing of coal fines in the coal-feed system lock vessel. These problems did not occur in TC15. The PRB average percent below 45 μm was 8.1 percent with a standard deviation of 3.2 percent, lower than typical PRB test runs. The majority of the coal samples had fines percentages around 8 percent. The values spiked occasionally to over 10 percent. Even during these periods, the FD0210 lock vessel did not experience any packing.

The TC15 standpipe solids particle sizes are given in [Figure 3.1-25](#). The PRB standpipe solids particle sizes remained constant as the run progressed due to the numerous sand additions in TC15. The average particle size of the standpipe solids was around 131 μm MMD and 151 μm SMD.

The percent of gasifier solids greater than 600 μm particles and the percent less than 45 μm are plotted on [Figure 3.1-26](#). During TC15 operation, the gasifier solids had an average coarse particle (greater than 600 μm) content of above 1.5 percent and an average fines content (less than 45 μm) of around 1.1 percent. The addition of sand kept particle size consistent for most of the test run, but increased to over 6 percent at the end of the test run.

For some of the previous test campaigns, the gasifier circulating solids achieved a steady particle size, typically between 165 and 205 μm SMD as shown in [Table 3.1-10](#). By comparison, the start-up sand is around 150 μm SMD. For tests that reached steady state, the standpipe particle size slowly increased asymptotically to reach the steady state value. Due to the short duration of the test run and the numerous sand additions, the standpipe solids never reached a steady state composition. The TC15 maximum standpipe particle size was lower than those of the previous PRB test campaigns as shown in [Table 3.1-10](#).

The particle sizes of the loop seal solids are shown in [Figure 3.1-27](#). Both the SMD and the MMD of the loop seal solids varied widely. The SMD ranged from 81.2 to 132.3 μm , while the MMD varied from 44.9 to 129.0 μm .

[Figure 3.1-28](#) plots the SMD and MMD for the PCD fines sampled from the fines transport systems, as well as, the seven in situ samples collected at the PCD inlet. Only two of the five in situ solids particle sizes agreed with the particle size of the solids collected from the hoppers. Sand addition is likely the reason for the disparity. In order to collect a representative sample of gasification ash, most of the in situ samples were collected during periods when sand was not being added to the gasifier. The hopper samples were often taken at times when sand had been added.

The PCD fines SMD started TC15 around 15 μm , and ranged from 7 to 17 μm for the remainder of the run. The MMD was about 10 μm larger than the SMD for most of the samples and followed the same trends as the SMD particle size. The TC15 PCD fines particle size was consistent with the particle size of previous PRB test campaigns as shown in [Table 3.1-10](#). The

particle size was lower than it was in TC14, indicating that the cyclone performance had improved.

Particle Size Comparison

Figure 3.1-29 plots the solids SMD particle sizes, including the coal, standpipe solids, loop seal solids, and PCD fines from fines transport systems. The coal fed to the Transport Gasifier averaged 216 μm SMD, slightly higher than the standpipe solids at 151 μm . The loop seal solids averaging 86 μm were always more coarse than the PCD solids (at an average of under 12 μm), but less coarse than the standpipe solids.

Standpipe and PCD Fines Bulk Densities

The standpipe, loop seal, PCD in situ, and FD0540 fines bulk densities are given in Figure 3.1-30. As previously mentioned, the standpipe solids bulk density remained constant due to the frequent sand additions that were necessary. The standpipe solids bulk density averaged 83 lbs/ft^3 , a value close to that of sand. Falling between the standpipe and PCD bulk density data, the data for the loop seal solids averaged 64 lbs/ft^3 , but varied considerably more than the standpipe solids data. Since the disengager captured most of the sand particles, the loop seal solids varied more with the density of the coal feed. The PCD solids had the lowest average values, around 20 lbs/ft^3 for the material removed from the fines systems. All of the in situ PCD data points were in the general range of the fines systems sample data. Of the three different sample locations, the PCD solids bulk density was the least dependent on sand addition and most dependent on the density of the coal feed.

The minimum standpipe solids densities for past PSDF gasification test campaigns are shown in Table 3.1-10.

3.1.5 Carbon Conversion

Carbon conversion is defined as the percent of fuel carbon that is gasified to CO , CO_2 , CH_4 , C_2H_6 , and higher hydrocarbons versus the amount of carbon that is rejected by the gasifier with the PCD and gasifier solids. In a typical flow sheet for an integrated gasification combine cycle (IGCC), the rejected carbon from the gasifier or PCD is burned in a combustor or sent to a landfill for disposal.

The carbon conversion can be calculated several different ways, the most accurate method divides the carbon content in the syngas by the total carbon exiting the gasifier (from both solid and gas streams). Table 3.1-11 gives the carbon conversions for the test periods, while Figure 3.1-31 shows the carbon conversion versus time. The carbon conversion ranged from 91 to 98 percent during the test run. The mode of operation (air or oxygen) did not appear to affect the carbon conversion. Note that the two highest carbon conversions were during periods of short duration (1.5 hours). TC15-01 was only 16 hours after startup. Neglecting these two periods (TC15-01 and TC15-16), all of the other operating periods were less than 96-percent carbon conversion. The carbon conversion was lower during periods in which the PCD solids rate was higher. During these periods, the solids collection system performed poorly, and more

fine carbon particles exited the gasifier without being recycled, thus lowering the carbon conversion.

The average carbon conversions of PRB, Hiawatha bituminous, Falkirk lignite coal, and Freedom lignite are compared in [Figure 3.1-32](#) for air and oxygen operation. This data came from Test Runs TC06 through TC13 and TC15. The graph does not include TC14 data, since the poor performance of the solids collection systems caused abnormally low carbon conversion. The low temperature Freedom lignite carbon conversion data is plotted separately from the high temperature Freedom lignite carbon conversion data to illustrate that significantly lower temperatures adversely affect the carbon conversion. For all fuels, air-blown operation yielded a slightly higher carbon conversion than oxygen-blown operation. The reason for the lower carbon conversion was because the oxygen system at the PSDF only supplied oxygen to the LMZ, causing the temperatures to be higher in that region and cooler through the rest of the gasifier. The cooler temperatures in the UMZ and riser kept the carbon conversion artificially low. Future tests with oxygen to the UMZ should improve the carbon conversion during oxygen-blown operations.

Falkirk lignite had the highest average carbon conversion of the four coals tested. PRB and Freedom lignite had about the same average carbon conversion, while Hiawatha bituminous had the lowest average carbon conversion. Although the data in [Figure 3.1-32](#) shows general trends in carbon conversion over test runs, the values obtained are the result of operating over a small range of conditions for all fuels except PRB coal.

3.1.6 Gasification Efficiencies

Gasification efficiency is defined as the percentage of the entering coal energy that is converted to potentially useful syngas energy. Two types of gasification efficiencies have been defined: the cold gasification efficiency and the hot gasification efficiency. The cold gasification efficiency is the amount of energy feed that is available to a gas turbine as syngas latent heat. The hot gasification efficiency is the percentage of total energy feed that is available to produce electricity. The total energy to produce electricity includes the syngas latent heat recovered in a gas turbine plus the sensible heat recovered in a steam turbine.

The cold gas efficiency is a measurement of the amount of latent heat in the coal feed to the gasifier that exits as syngas latent heat. The cold gasification efficiency is plotted in [Figure 3.1-33](#) and is listed in [Table 3.1-11](#). Note that in previous run reports (TC06-TC14), the gasification efficiency was defined as the amount of latent heat in the syngas divided by the total amount of heat entering the gasifier (including the air and steam sensible heats). This report includes the modified calculation in order to more closely conform to industry standards for determining cold gasification efficiency. Thus, the values in previous test run reports are artificially lower than those in this report. During TC15, the cold gasification efficiencies were as high as 64.3 percent during the test run. During the air-blown operating periods, the cold gas efficiencies ranged from 35.7 to 58.6 percent. The oxygen-blown periods had cold gas efficiencies of between 60.6 and 64.3 percent. Since less nitrogen had to be heated to gasifier temperature in oxygen-blown mode, the efficiencies tend to be higher.

The hot gasification efficiency assumes that the sensible heat of the syngas can be recovered in a heat recovery steam generator, so the hot gasification efficiency is always higher than the cold

gasification efficiency. The hot gasification efficiency is the latent and the sensible heat of the syngas exiting the gasifier divided by the total amount of energy entering the gasifier, including the latent heat of the coal and the sensible heats of the air and steam. The hot gasification efficiency is plotted in [Figure 3.1-34](#) and shown in [Table 3.1-11](#). The efficiency boundary for the values found in [Table 3.1-11](#) and [Figure 3.1-34](#) is the gasifier itself, not including any downstream equipment.

The air-blown hot gas efficiencies were between 70.4 and 86.6 percent, with the periods of lowest efficiency occurring either during the periods of high PCD solids carryover and low carbon conversion or low coal-feed rate.

The two main sources of efficiency losses are the gasifier heat loss and the latent heat of the PCD solids. The gasifier heat loss of 3.5 MBtu/hr was about 10 percent of the feed energy, while the total energy of the PCD solids was from 2 to 8 percent of the feed energy (the higher numbers occurring during the periods of low coal-feed rate).

A commercial gasifier will be more efficient than the PSDF gasifier due to the use of recycled gas and lower heat losses. The heat loss as a percentage of energy feed will be much smaller in a commercially sized gasifier. While the Transport Gasifier does not recover the latent heat of the PCD solids, this latent heat could be recovered in a combustor. The total enthalpy of the PCD solids can be decreased by decreasing both the PCD solids carbon content (heating value) and the PCD solids rate (by improving solids collection efficiency).

Gasification efficiencies can be calculated from the commercially projected gas heating values and adjusted flow rates that were determined when calculating the projected heating value. The commercially projected cold gasification efficiencies are listed on [Table 3.1-11](#) for all of the operating periods. The projected hot gasification efficiencies are not shown because the commercial projections do not increase the hot gasification efficiency because the deleted nitrogen lowers the syngas sensible heat and increases the syngas latent heat. Both changes effectively cancel each other.

The corrected efficiencies are calculated assuming an adiabatic gasifier, since zero heat loss was one of the assumptions in determining the corrected LHV in Section 3.1.3. The corrected cold gas efficiencies were from 64.9 to 76.6 percent in air-blown mode and between 77.8 and 84.3 percent in oxygen-blown mode. When projected for a commercial plant, the cold gas efficiency for air-blown mode averaged 72.4 percent. Oxygen-blown projections averaged 79.4 percent. The commercially projected efficiencies were higher than the observed cold gasification efficiencies by about 19 percent in air-blown mode and 18 percent in oxygen-blown mode. The use of recycled gas and the lower heat losses in the commercial projection are the main factors that increase the efficiency.

Table 3.1-1

Operating Periods

Operating Period	Start Time	End Time	Duration Hours	Operating Period	
				Average Time	Relative Hours
TC15-1	4/20/2004 14:30	4/20/2004 16:00	1:30	4/20/2004 15:15	16
TC15-2	4/21/2004 5:00	4/21/2004 7:45	2:45	4/21/2004 6:22	31
TC15-3	4/21/2004 13:00	4/21/2004 16:30	3:30	4/21/2004 14:45	40
TC15-4	4/22/2004 0:30	4/22/2004 7:00	6:30	4/22/2004 3:45	53
TC15-5	4/22/2004 7:15	4/22/2004 9:00	1:45	4/22/2004 8:07	57
TC15-6	4/22/2004 23:00	4/23/2004 3:15	4:15	4/23/2004 1:07	74
TC15-7	4/23/2004 5:15	4/23/2004 8:00	2:45	4/23/2004 6:37	80
TC15-8	4/23/2004 10:15	4/23/2004 13:00	2:45	4/23/2004 11:37	85
TC15-9	4/23/2004 15:00	4/23/2004 17:00	2:00	4/23/2004 16:00	89
TC15-10	4/23/2004 21:30	4/23/2004 23:45	2:15	4/23/2004 22:37	96
TC15-11	4/24/2004 19:00	4/24/2004 21:45	2:45	4/24/2004 20:22	117
TC15-12	4/25/2004 2:00	4/25/2004 4:15	2:15	4/25/2004 3:07	124
TC15-13	4/25/2004 6:00	4/25/2004 9:00	3:00	4/25/2004 7:30	128
TC15-14	4/25/2004 15:00	4/25/2004 20:15	5:15	4/25/2004 17:37	139
TC15-15	4/25/2004 23:30	4/26/2004 1:30	2:00	4/26/2004 0:30	145
TC15-16	4/27/2004 5:15	4/27/2004 6:45	1:30	4/27/2004 6:00	161
TC15-17	4/29/2004 11:00	4/29/2004 12:15	1:15	4/29/2004 11:37	201

Table 3.1-2

Operating Conditions

Operating Periods	Average Relative Hours	Riser Exit Temperature TI352 °F	Pressure PI287 psig	Coal Rate ² lb/hr	Air Rate lb/hr	Oxygen Rate lb/hr	Syngas Rate lb/hr	Steam Rate ³ lb/hr	Nitrogen Rate lb/hr	PCD Inlet Temperature TI458 °F	PCD Solids Rate lb/hr
TC15-1	16	1,682	174	3,600	10,900	0	20,400	1,000	6,200	812	210
TC15-2	31	1,686	200	4,000	12,400	0	22,200	900	6,700	827	340
TC15-3	40	1,673	214	4,500	13,200	0	22,900	800	6,400	819	340
TC15-4	53	1,633	214	4,500	12,600	0	22,400	800	6,500	801	260
TC15-5	57	1,654	214	4,500	12,900	0	22,600	800	6,400	808	240
TC15-6	74	1,668	214	4,800	13,200	0	22,000	200	5,900	771	310
TC15-7	80	1,753	150	2,100	8,800	0	15,300	500	5,700	727	340
TC15-8	85	1,720	134	3,200	800	2,200	12,400	1,900	5,500	689	370
TC15-9	89	1,732	134	4,100	1,000	2,600	14,700	2,700	5,800	715	370
TC15-10	96	1,717	134	4,200	900	2,500	14,700	2,700	5,800	718	370
TC15-11	117	1,684	144	4,200	1,000	2,600	15,700	3,400	6,000	735	370
TC15-12	124	1,686	144	4,200	800	2,600	15,700	3,300	6,000	736	370
TC15-13	128	1,684	144	4,200	700	2,500	15,300	3,000	6,000	725	370
TC15-14	139	1,696	150	4,100	800	2,500	14,800	2,700	6,100	736	370
TC15-15	145	1,719	150	3,900	600	2,500	14,800	2,700	6,100	783	370
TC15-16	161	1,537	111	3,200	200	2,000	13,900	2,900	5,600	795	150
TC15-17	201	1,409	127	1,500	6,900	0	14,500	700	6,100	888	120

Notes:

1. TC15-1 to TC15-7 and TC15-17 were air blown; TC15-8 to TC15-16 were oxygen blown.
2. TC15-1 to TC15-15 coal rates by weigh cells; TC15-16 to TC15-17 coal rates by syngas combustor carbon balance.
3. All steam rates by hydrogen balance

Table 3.1-3

Wet Gas Compositions, Molecular Weight, and Heating Value

Operating Period ²	Average Relative Hour	H ₂ O ³ Mole %	CO Mole %	H ₂ Mole %	CO ₂ Mole %	CH ₄ Mole %	C ₂ H ₆ Mole %	Argon Mole %	N ₂ Mole %	Total Mole %	Syngas LHV Btu/SCF	Syngas TRS ¹ ppm	Overall %O ₂ %	CO/CO ₂ Ratio mol/mol	Syngas MW lb/Mole
TC15-1	16	10.7	7.7	6.7	9.1	1.5	0.0	0.5	63.7	100.0	57	340	12.1	0.84	26.5
TC15-2	31	9.6	8.4	6.8	8.9	1.6	0.0	0.5	64.1	100.0	61	343	12.5	0.94	26.6
TC15-3	40	9.1	9.3	7.2	9.0	1.9	0.0	0.5	63.0	100.0	67	391	13.2	1.03	26.5
TC15-4	53	9.8	8.4	7.2	9.4	1.9	0.0	0.5	62.9	100.0	64	455	12.8	0.90	26.5
TC15-5	57	10.0	8.4	7.1	9.2	1.8	0.0	0.5	63.1	100.0	63	437	13.0	0.90	26.5
TC15-6	74	7.0	11.4	7.3	8.4	1.9	0.0	0.5	63.5	100.0	74	431	14.0	1.35	26.6
TC15-7	80	9.5	6.3	4.8	8.4	0.6	0.0	0.5	69.9	100.0	39	394	11.9	0.75	27.1
TC15-8	85	21.6	10.9	12.8	13.0	2.2	0.0	0.0	39.4	100.0	91	732	18.7	0.83	24.4
TC15-9	89	24.0	11.5	14.0	13.8	2.5	0.0	0.1	34.2	100.0	98	933	18.9	0.83	23.9
TC15-10	96	25.8	10.5	13.8	14.0	2.3	0.0	0.0	33.6	100.0	92	908	18.2	0.75	23.8
TC15-11	117	28.5	8.7	13.1	14.1	2.4	0.0	0.0	33.3	100.0	86	887	17.1	0.61	23.7
TC15-12	124	27.4	9.1	13.3	14.0	2.5	0.0	0.0	33.6	100.0	89	901	17.2	0.65	23.8
TC15-13	128	26.1	9.6	13.4	14.0	2.5	0.0	0.0	34.3	100.0	91	892	17.5	0.69	23.9
TC15-14	139	25.1	9.9	13.3	13.7	2.3	0.0	0.0	35.7	100.0	90	853	18.0	0.72	24.0
TC15-15	145	24.6	10.5	13.6	13.4	1.9	0.0	0.0	35.9	100.0	89	801	18.0	0.78	23.9
TC15-16	161	26.9	6.7	11.8	14.7	2.8	0.1	0.1	37.0	100.0	80	885	15.1	0.46	24.3
TC15-17	201	10.1	2.3	3.1	8.9	1.0	0.0	0.4	74.1	100.0	26	298	10.1	0.26	27.6

Notes:

1. Syngas total reduced sulfur (TRS) estimated from syngas combustor SO₂ analyzer data.
2. TC15-1 to TC15-7 and TC15-17 were air blown; TC15-8 to TC15-16 were oxygen blown.
3. The H₂O concentration was estimated using a correlation between the in-situ samples and gasifier temperature.

Table 3.1-4

Syngas Sulfur and Ammonia Concentrations

Operating Period ¹	Average Relative Hour	Wet AI419J H ₂ S ppm	Syngas Total Reduced Sulfur ² ppm	Wet AI475Q Ammonia ppm
TC15-1	16	13	304	1,401
TC15-2	31	3	310	1,548
TC15-3	40	75	355	1,891
TC15-4	53	150	410	1,985
TC15-5	57	159	394	1,883
TC15-6	74	166	401	1,793
TC15-7	80	157	356	752
TC15-8	85	126	574	2,206
TC15-9	89	333	709	2,563
TC15-10	96	366	673	2,478
TC15-11	117	441	635	2,264
TC15-12	124	451	654	2,344
TC15-13	128	463	659	2,355
TC15-14	139	460	639	1,987
TC15-15	145	453	604	1,779
TC15-16	161	223	647	3,988
TC15-17	201	70	268	1,159

Notes:

1. TC15-1 to TC15-7 and TC15-17 were air blown; TC15-8 to TC15-16 were oxygen blown.
2. Syngas total reduced sulfur (TRS) calculated from syngas combustor SO₂ analyzer data.

Table 3.1-5

Adjusted¹ Wet Syngas Compositions, Molecular Weight, and Projected Heating Value

Operating Period	Average Relative Hour	H ₂ O Mole %	CO Mole %	H ₂ Mole %	CO ₂ Mole %	CH ₄ Mole %	C ₂ H ₆ Mole %	Argon Mole %	N ₂ Mole %	Total Mole %	Syngas MW lb/lbmole	Overall %O ₂ %	Syngas CO/CO ₂ Ratio	Syngas LHV Btu/SCF
TC15-1	16	1.0	16.2	17.1	13.5	3.5	0.0	0.7	48.1	100.0	25.3	16.1	1.2	132
TC15-2	31	1.0	16.4	15.5	12.3	3.4	0.0	0.7	50.7	100.0	25.5	16.7	1.3	127
TC15-3	40	1.0	16.9	15.0	11.9	3.6	0.0	0.7	50.9	100.0	25.6	17.2	1.4	129
TC15-4	53	1.0	15.9	15.6	12.9	3.8	0.0	0.7	50.2	100.0	25.6	17.3	1.2	129
TC15-5	57	1.0	15.6	15.2	12.7	3.6	0.0	0.7	51.3	100.0	25.6	17.2	1.2	125
TC15-6	74	1.0	19.8	14.2	10.0	3.6	0.0	0.6	50.9	100.0	25.5	19.0	2.0	135
TC15-7	80	1.0	15.2	15.3	11.4	1.6	0.0	0.6	54.8	100.0	25.7	16.3	1.3	106
TC15-8	85	1.0	34.3	37.2	20.7	6.8	0.0	0.0	0.0	100.0	20.7	29.6	1.7	275
TC15-9	89	1.0	32.4	36.7	23.2	6.8	0.0	0.0	0.0	100.0	21.3	28.6	1.4	267
TC15-10	96	1.0	30.8	37.3	24.6	6.3	0.0	0.0	0.0	100.0	21.4	29.7	1.3	260
TC15-11	117	1.0	27.4	37.4	27.1	7.1	0.0	0.0	0.0	100.0	21.7	25.3	1.0	256
TC15-12	124	1.0	28.7	37.0	25.9	7.4	0.0	0.0	0.0	100.0	21.5	25.7	1.1	262
TC15-13	128	1.0	29.7	36.8	25.1	7.3	0.0	0.0	0.0	100.0	21.5	27.4	1.2	264
TC15-14	139	1.0	30.4	37.3	24.5	6.8	0.0	0.0	0.0	100.0	21.3	28.5	1.2	263
TC15-15	145	1.0	32.2	38.0	23.2	5.7	0.0	0.0	0.0	100.0	21.1	28.5	1.4	260
TC15-16	161	1.0	27.5	34.6	27.3	9.4	0.2	0.0	0.0	100.0	22.1	25.9	1.0	272
TC15-17	201	1.0	8.4	16.8	17.8	4.7	0.1	0.7	50.6	100.0	25.9	12.5	0.5	118

Notes:

- Adjustments are based on the following assumptions: that only air nitrogen is in the syngas, the gasifier is adiabatic, and syngas is at the turbine inlet after the syngas cleanup processes.
- TC15-1 to TC15-7 and TC15-17 were air blown; TC15-8 to TC15-16 were oxygen blown.

Table 3.1-6

Coal Analysis

	Powder River Basin	
	Average Value ¹	Standard Deviation
Moisture, wt%	21.27	2.35
Carbon, wt%	55.11	1.68
Hydrogen ² , wt%	3.46	0.14
Nitrogen, wt%	0.73	0.03
Oxygen, wt%	14.12	0.74
Sulfur, wt%	0.24	0.02
Ash, wt%	5.08	0.30
Volatiles, wt%	34.24	1.03
Fixed Carbon, wt%	39.41	1.89
Higher Heating Value, Btu/lb	9,143	237
Lower Heating Value, Btu/lb	8,823	229
CaO, wt %	0.86	0.05
SiO ₂ , wt %	1.69	0.12
Al ₂ O ₃ , wt %	0.80	0.04
MgO, wt %	0.27	0.03
Na ₂ O, wt %	0.80	0.04
Fe ₂ O ₃ , wt %	0.26	0.01
Na, wt % in ash	1.13	0.24
Ca/S, mole/mole	2.82	0.36

Notes:

1. All analyses are as sampled at FD0210.
2. Hydrogen in coal is reported separately from hydrogen in moisture.

Table 3.1-7

Standpipe Solids Analysis

Sample Number ^{1,2}	Sample Date & Time	Sample Run Time Hours	SiO ₂ Wt. %	Al ₂ O ₃ Wt. %	FeO Wt. %	Other Inerts ³ Wt. %	CaCO ₃ Wt. %	CaS Wt. %	CaO Wt. %	MgO Wt. %	Organic Carbon Wt. %	Total Wt. %
AB14738	4/20/2004 10:00	11	90.5	2.6	0.5	2.0	1.4	0.1	0.6	0.3	1.2	99.4
AB14797	4/23/2004 19:00	92	79.3	4.7	1.4	3.8	3.4	0.1	3.7	1.2	1.4	99.0
AB14799	4/24/2004 10:00	107	84.6	4.3	1.2	2.5	2.9	0.1	3.1	1.1	0.4	100.2
AB14800	4/25/2004 10:00	131	84.2	4.4	1.3	2.7	3.2	0.1	3.3	1.1	0.4	100.7
AB14801	4/25/2004 20:00	141	79.8	5.3	1.6	2.9	2.1	0.1	5.2	1.3	0.4	98.7
AB14845	4/29/2004 9:30	199	84.1	4.1	1.4	2.2	1.9	0.1	3.1	1.0	2.2	100.1

Notes:

1. Samples on April 20 and April 29 were taken during air blown mode; Samples taken from April 23 to April 25 were taken during oxygen blown mode
2. April 20 sample taken from standpipe; April 23 to 29 samples taken from FD0510.
3. Other inerts consist of P₂O₅, Na₂O, K₂O, BaO, & TiO₂

Table 3.1-8

Loop Seal Solids Sample Analysis

Sample ² Number	Sample Date & Time	Sample Run Time Hours	SiO ₂ Wt. %	Al ₂ O ₃ Wt. %	FeO Wt. %	Other Inerts ¹ Wt. %	CaCO ₃ Wt. %	CaS Wt. %	CaO Wt. %	MgO Wt. %	Organic Carbon Wt. %	Total Wt. %
AB14737	4/20/2004 10:00	11	86.3	4.0	0.8	2.7	2.1	0.0	1.4	0.6	1.0	98.8
AB14747	4/21/2004 10:00	35	74.7	5.4	1.4	3.7	3.2	0.0	4.2	1.6	2.1	96.3
AB14778	4/23/2004 10:00	83	60.2	9.1	2.6	3.6	4.2	0.2	8.4	2.4	9.4	100.1
AB14846	4/29/2004 10:00	199	40.6	2.6	0.7	1.2	3.1	0.1	0.6	0.5	52.7	102.2

Notes:

1. Other inerts consist of P₂O₅, Na₂O, K₂O, BaO, & TiO₂
2. April 20, April 21, and April 29 samples taken during air blown operation; April 23 sample taken during oxygen blown operation.

Table 3.1-9

PCD Fines Solids From FD0520 and FD0540 Analysis

Sample ^{2,3} Number	Sample Date & Time	Sample Run Time Hours	SiO ₂ Wt. %	Al ₂ O ₃ Wt. %	FeO Wt. %	Other Inerts ¹ Wt. %	CaCO ₃ Wt. %	CaS Wt. %	CaO Wt. %	MgO Wt. %	Organic C (C-CO ₂) Wt. %	Total Wt. %	HHV Btu/lb.	LHV Btu/lb.
AB14740	4/20/2004 14:00	15	58.5	8.0	2.4	3.5	3.4	0.3	8.5	2.6	9.7	96.8	1,722	1,716
AB14741	4/20/2004 22:00	23	45.0	6.4	1.7	2.8	4.7	0.6	4.7	2.0	40.4	108.3	4,768	4,715
AB14742	4/21/2004 2:00	27	30.2	6.8	1.8	2.5	5.5	0.8	4.5	2.1	49.3	103.4	6,707	6,634
AB14748	4/21/2004 10:00	35	40.3	7.5	2.2	3.0	4.5	0.8	7.0	2.5	32.3	100.1	4,826	4,785
AB14749	4/21/2004 14:00	39	30.7	7.6	0.9	3.2	4.8	0.9	8.7	2.7	39.3	98.8	5,806	5,755
AB14761	4/21/2004 18:00	43	23.9	7.2	2.4	2.7	5.4	0.9	7.1	2.6	48.8	101.0	7,161	7,099
AB14763	4/22/2004 2:00	51	20.5	7.6	2.6	2.5	5.4	0.9	7.8	2.7	48.2	98.2	7,391	7,325
AB14764	4/22/2004 6:00	55	19.7	7.3	2.2	2.3	5.7	0.8	6.2	2.5	50.5	97.4	7,834	7,772
AB14777	4/23/2004 10:00	83	41.3	9.1	2.7	3.0	5.6	0.7	8.3	2.7	25.4	98.7	3,962	3,944
AB14802	4/23/2004 18:00	91	35.8	9.8	3.5	3.7	7.4	0.4	10.6	3.4	26.0	100.7	3,896	3,871
AB14803	4/23/2004 22:00	95	30.6	9.9	3.1	3.0	6.9	0.4	9.1	3.2	34.3	100.5	5,093	5,055
AB14808	4/24/2004 18:00	115	27.0	10.0	3.5	3.0	7.5	0.3	10.5	3.7	33.5	99.0	5,228	5,189
AB14810	4/25/2004 2:00	123	31.0	6.7	2.1	2.6	6.4	0.8	5.3	3.0	40.3	98.1	6,213	6,159
AB14811	4/25/2004 6:00	127	29.7	9.0	2.7	3.1	4.0	1.2	9.2	3.1	36.0	98.0	5,472	5,426
AB14813	4/25/2004 14:00	135	24.6	9.0	2.8	2.9	4.0	1.1	9.3	3.0	40.8	97.5	6,215	6,165
AB14829	4/27/2004 18:00	173	30.2	10.8	3.2	3.3	4.6	1.2	11.1	3.6	29.3	97.4	4,369	4,334
AB14840	4/29/2004 2:00	191	52.9	8.4	2.4	4.1	5.0	0.4	6.2	2.1	17.5	99.0	2,709	2,692
AB14844	4/29/2004 10:00	199	27.2	5.5	1.5	2.4	5.5	0.5	2.9	1.5	49.4	96.5	7,772	7,689

Notes:

1. Other inerts consist of P₂O₅, Na₂O, K₂O, BaO, & TiO₂
2. Air blown samples from April 20 to 22 and April 29; oxygen blown samples from April 23 to 27.
3. April 20 to April 27 samples taken from FD0540; April 29 samples taken from FD0520.

Table 3.1-10

Historical Standpipe and PCD Fines

Test Campaign	Fuel	Standpipe				PCD Fines			
		Maximum	Steady	Minimum		Average	St. Dev.	Average	
		Particle Size	State	Part. Size	Bulk	Particle Size	Particle Size	Bulk	St. Dev.
		SMD	SMD	SMD	Density	SMD	SMD	Density	Bulk
		microns	microns	microns	lb/ft ³	microns	microns	lb/ft ³	Density lb/ft ³
TC06	Powder River Basin	204	165		80	10.8	1.1	24	4
TC07	Powder River Basin	191	175		80	10.2	1.1	28	8
TC07	Alabama Bituminous	232	none		66	16.2	3.2	32	7
TC08	Powder River Basin	250	205		77	13.1	3.2	25	7
TC09	Hiawatha Bituminous	233	180		76	15.7	4.6	29	12
TC10	Powder River Basin	280	none		76	10.7	3.6	23	7
TC11	Falkirk Lignite	200	200		75	12.3	2.4	36	3
TC12	Powder River Basin	300	none		76	9.8	2	18	6
TC13	Powder River Basin	165	165		81	10.4	1.4	18	4
TC13	Freedom Lignite Low Sodium	230	none		56	15.3	3.9	26	6
TC13	Freedom Lignite High Sodium, High Temp.	425	none		46	30.0	32.3	39	14
TC13	Freedom Lignite High Sodium, Low Temp.	457	none		67	13.9	2.3	26	5
TC14	Powder River Basin	220	none		84	18.7	14.6	27	14
TC15	Powder River Basin	156	none		79	10.7	1.3	20	4

Table 3.1-11

Carbon Conversion and Gasifier Efficiencies

	Average	Carbon	Efficiency		
			Raw		Projected ²
Operating	Relative	Conversion	Cold	Hot	Cold
Period ¹	Hours	%	%	%	%
TC15-1	16	98.1	56.5	86.6	76.6
TC15-2	31	93.2	55.9	84.2	72.4
TC15-3	40	93.9	58.6	85.3	73.5
TC15-4	53	93.9	57.8	84.7	73.4
TC15-5	57	94.4	57.5	85.1	73.2
TC15-6	74	95.3	61.3	86.4	75.9
TC15-7	80	91.4	41.3	75.1	64.9
TC15-8	85	94.4	60.6	82.6	80.3
TC15-9	89	95.6	64.3	86.0	80.7
TC15-10	96	94.0	62.1	84.4	78.7
TC15-11	117	93.9	61.3	84.3	78.0
TC15-12	124	93.4	61.8	84.2	78.2
TC15-13	128	93.7	62.2	84.3	78.8
TC15-14	139	92.8	61.0	83.2	77.9
TC15-15	145	92.8	60.6	83.1	77.8
TC15-16	161	97.4	62.7	85.3	84.3
TC15-17	201	93.2	35.7	70.4	69.5

Notes:

1. TC15-1 to TC15-7 and TC15-17 were air blown; TC15-8 to TC15-16 were oxygen blown.
2. Projection assumes that only air nitrogen in the syngas is from air and that the gasifier is adiabatic.

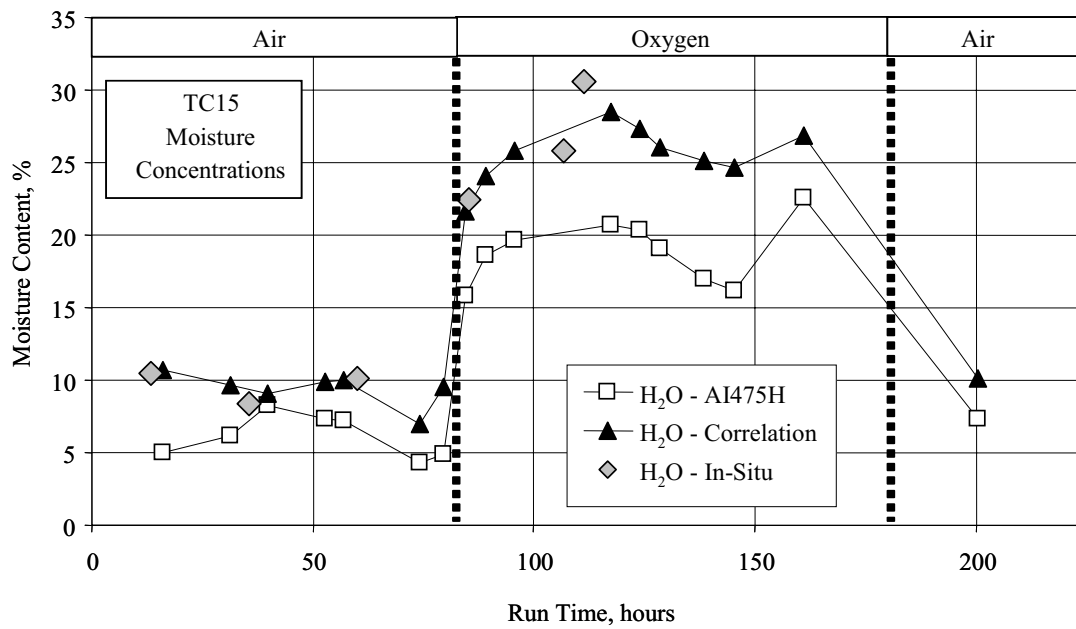


Figure 3.1-1 H₂O Data

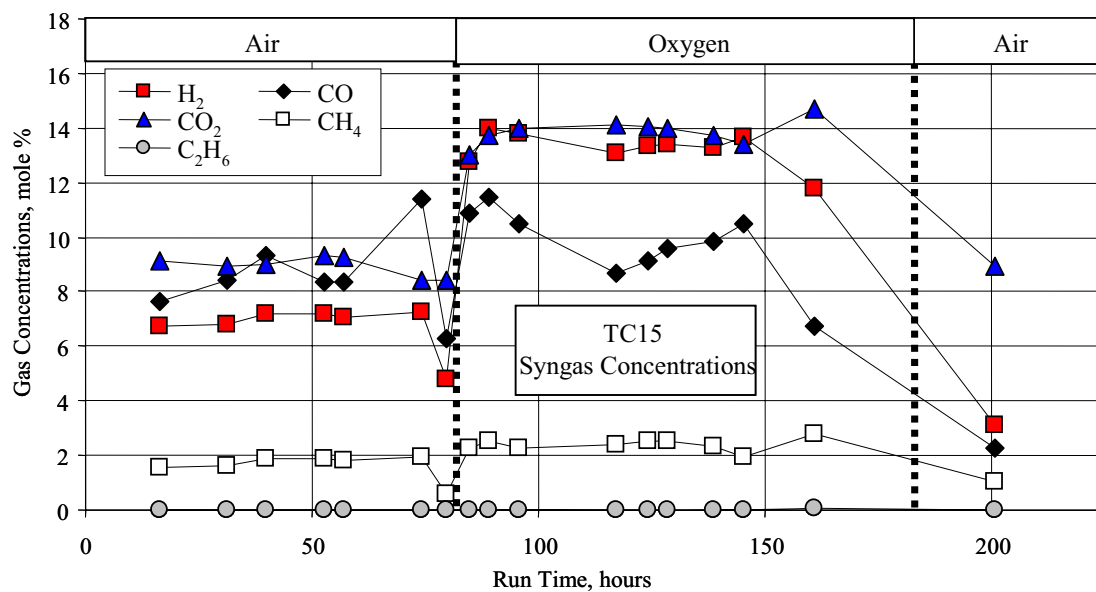


Figure 3.1-2 Wet Syngas Compositions

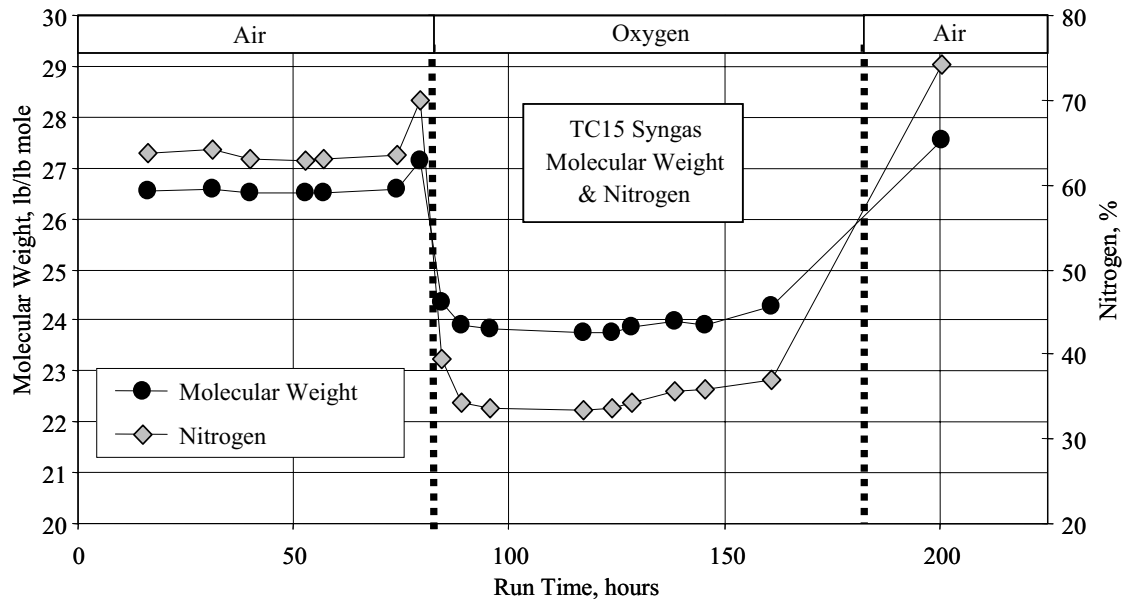


Figure 3.1-3 Wet Syngas Molecular Weight and Nitrogen Concentration

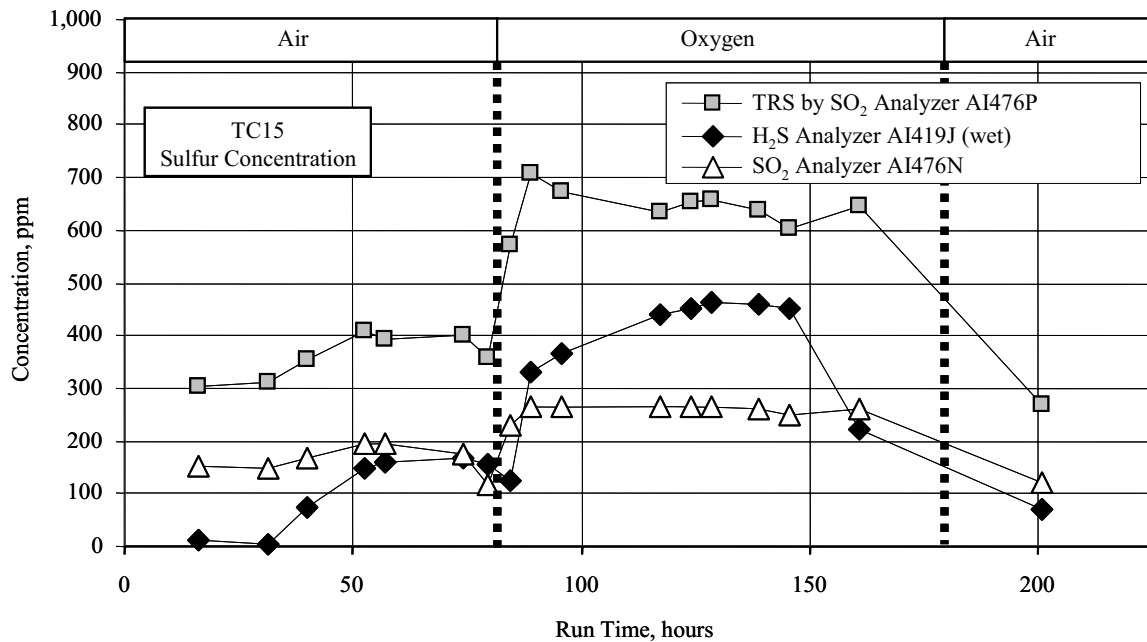


Figure 3.1-4 Sulfur Concentrations

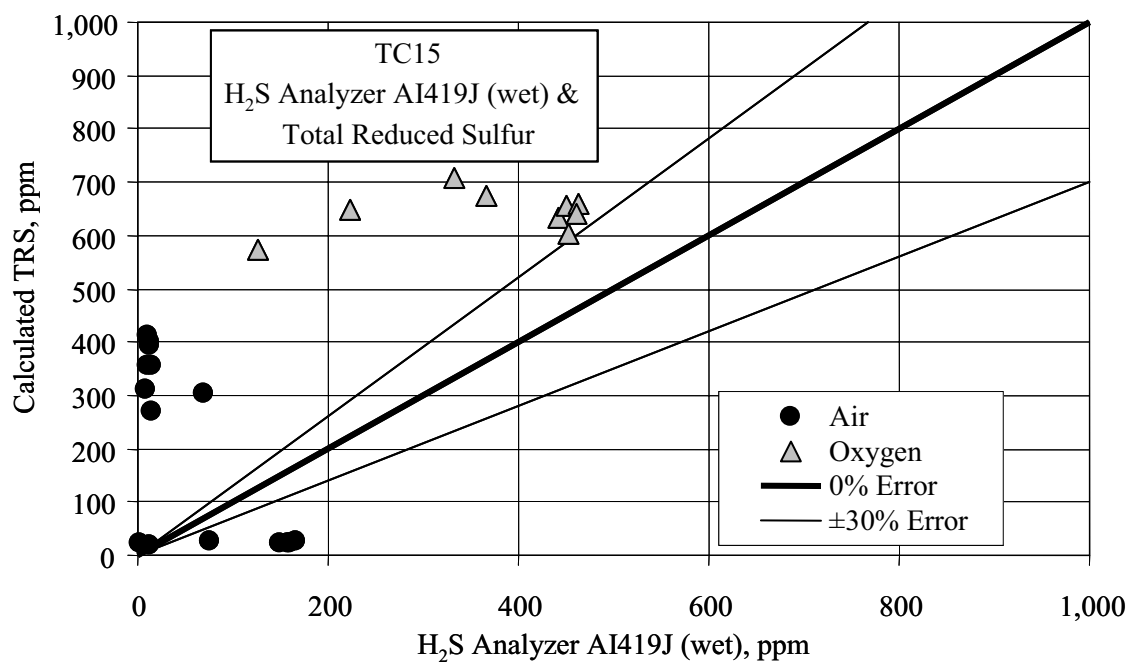


Figure 3.1-5 H₂S Analyzer AI419J and Total Reduced Sulfur

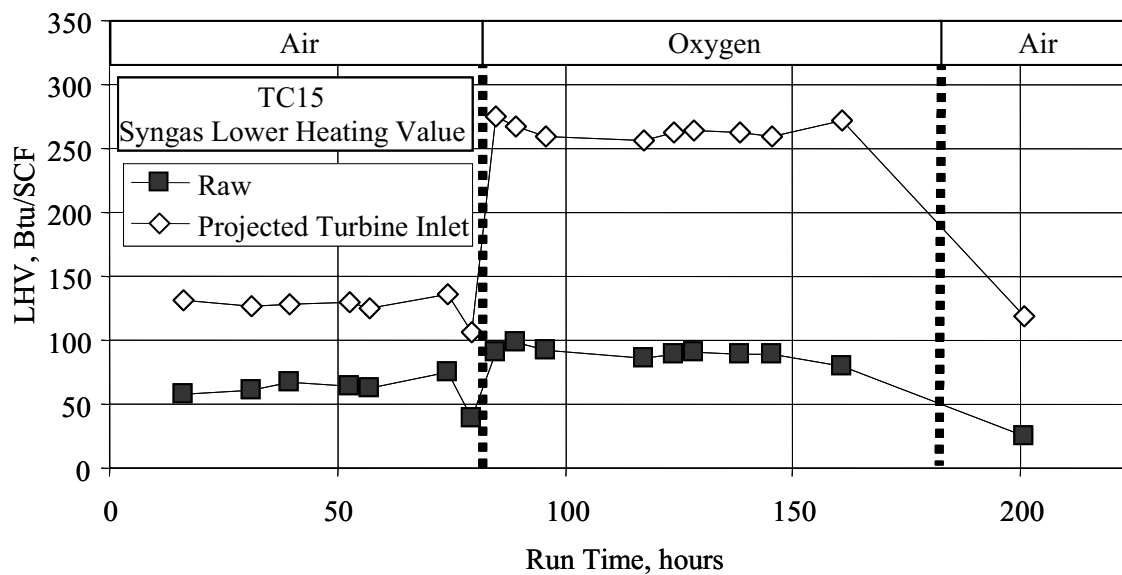


Figure 3.1-6 Syngas Lower Heating Values

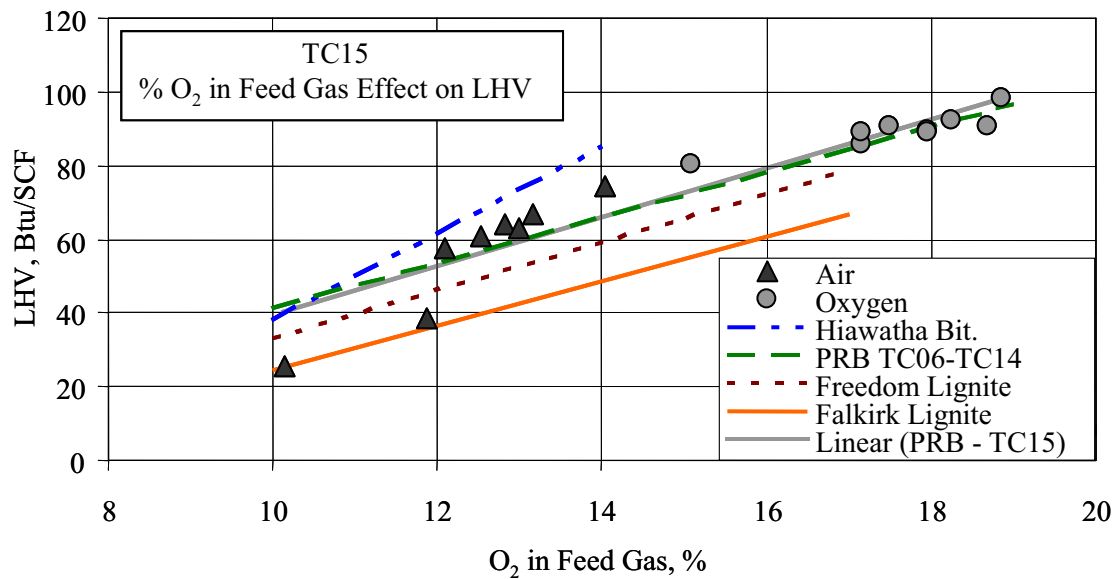


Figure 3.1-7 Raw Lower Heating Value and Overall Percent O₂

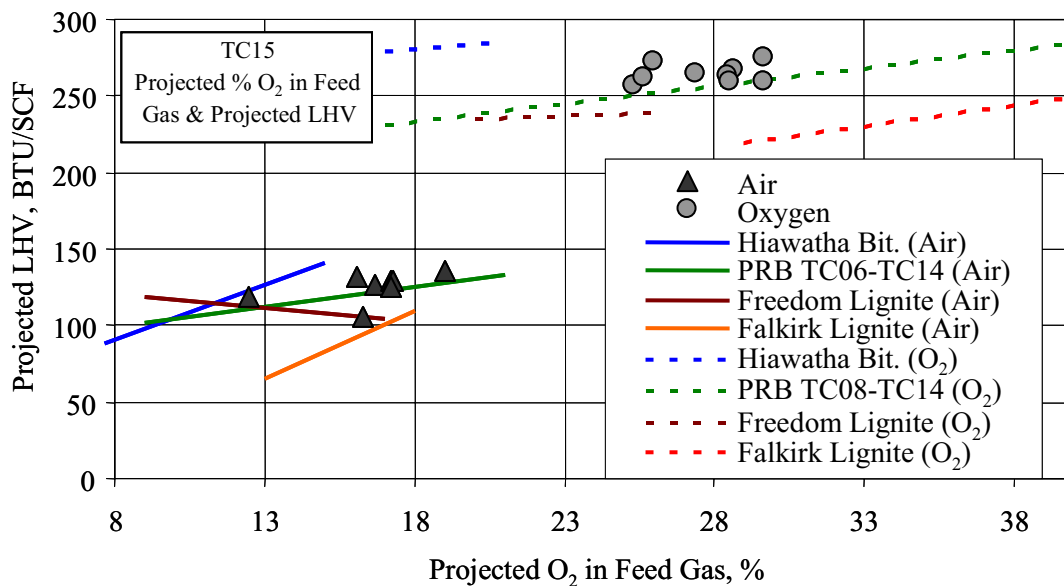


Figure 3.1-8 Projected LHV and Projected Overall Percent O₂

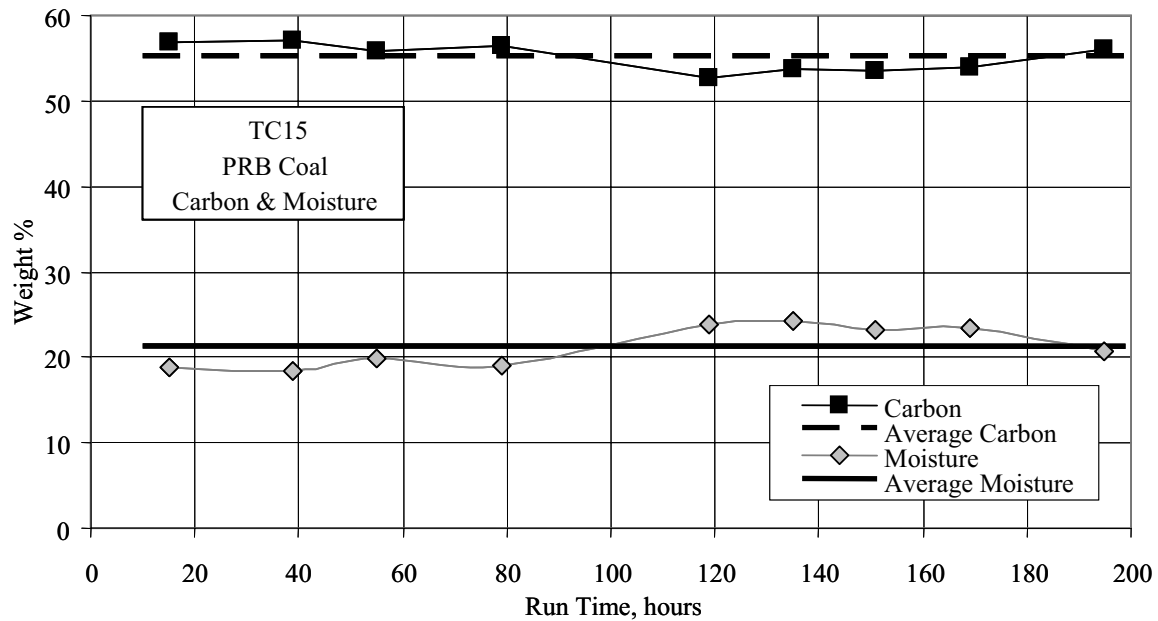


Figure 3.1-9 Coal Carbon and Moisture

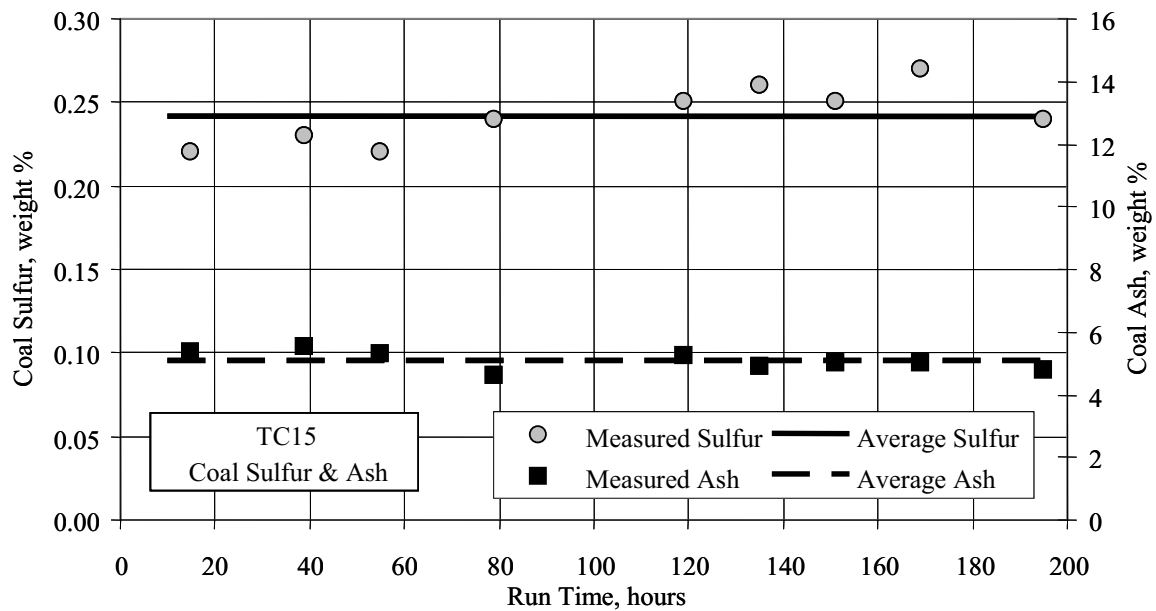


Figure 3.1-10 Coal Sulfur and Ash

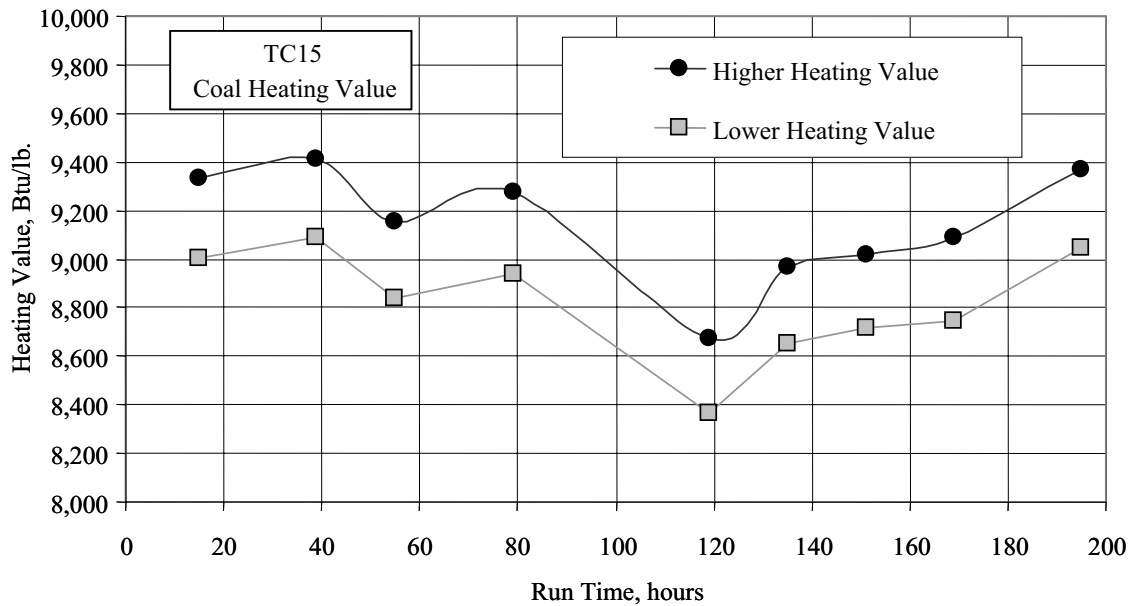


Figure 3.1-11 Coal Heating Value

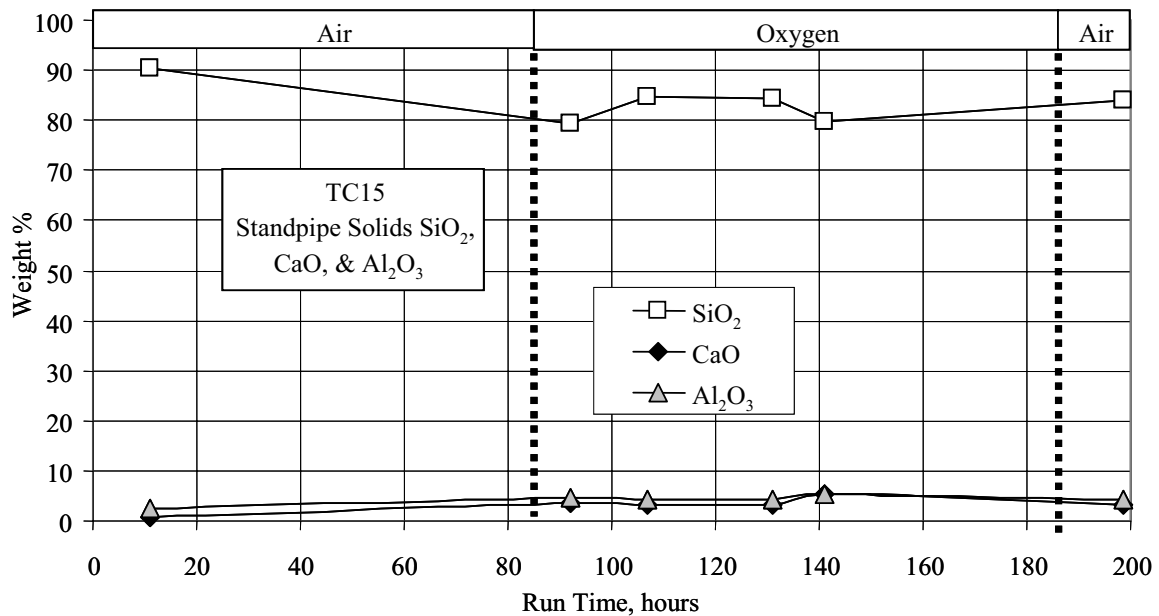


Figure 3.1-12 Standpipe Solids SiO_2 , CaO , and Al_2O_3

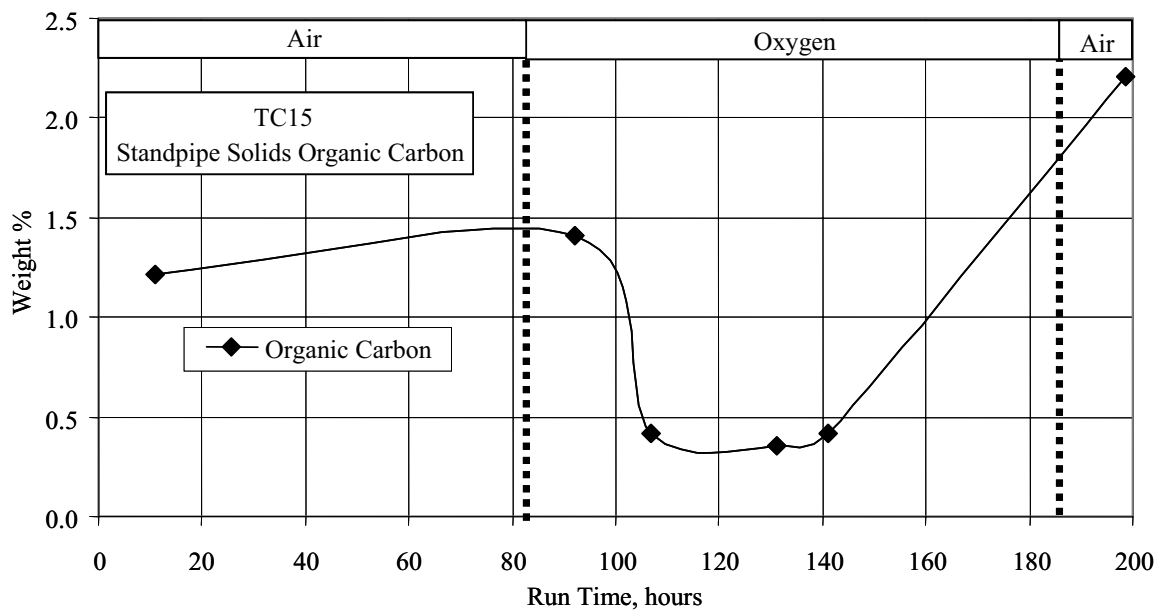


Figure 3.1-13 Standpipe Solids Organic Carbon

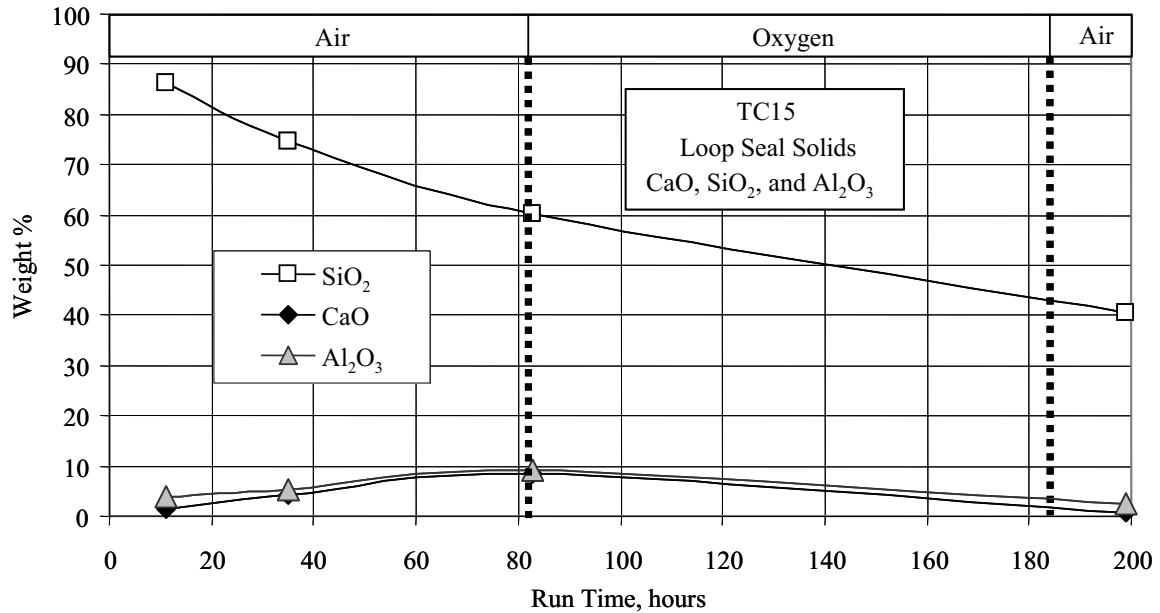


Figure 3.1-14 Loop Seal Solids SiO₂, CaO, and Al₂O₃

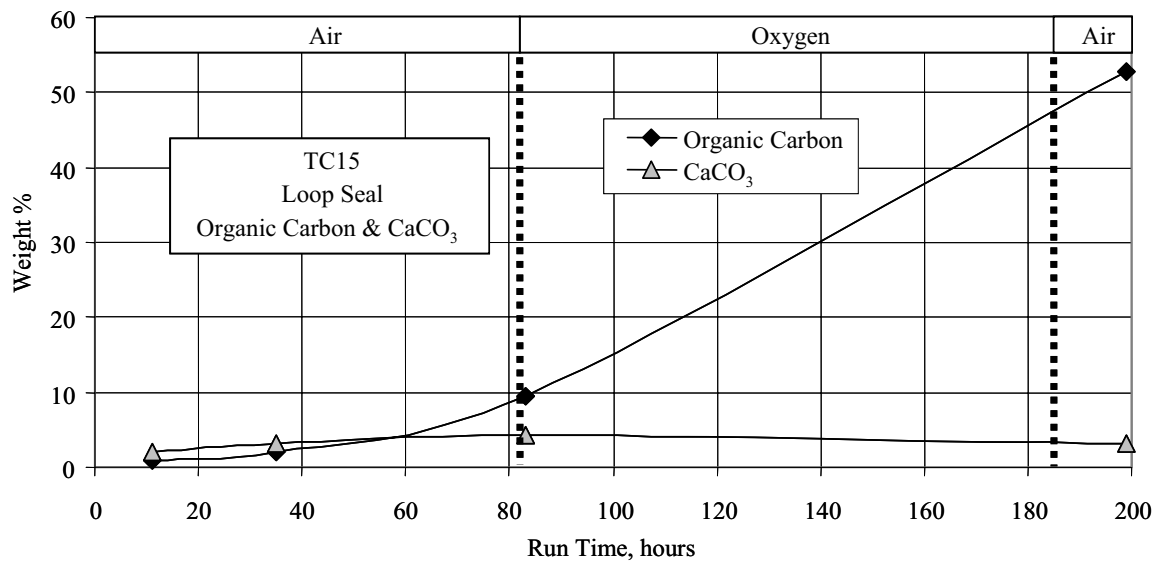


Figure 3.1-15 Loop Seal Solids Organic Carbon and CaCO₃

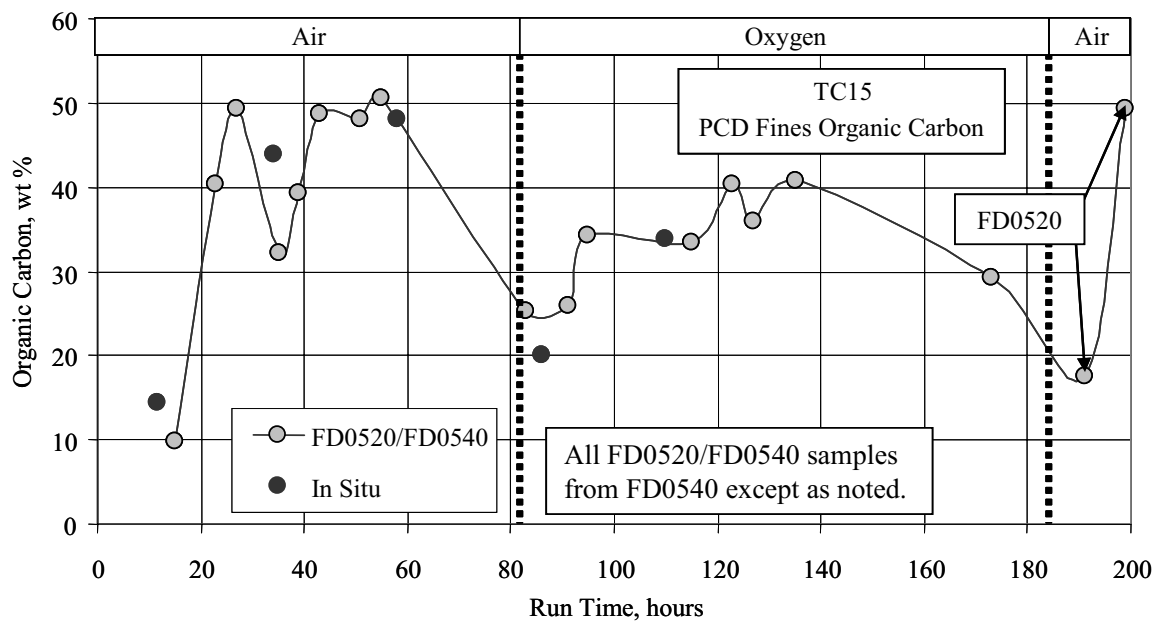


Figure 3.1-16 PCD Fines Organic Carbon

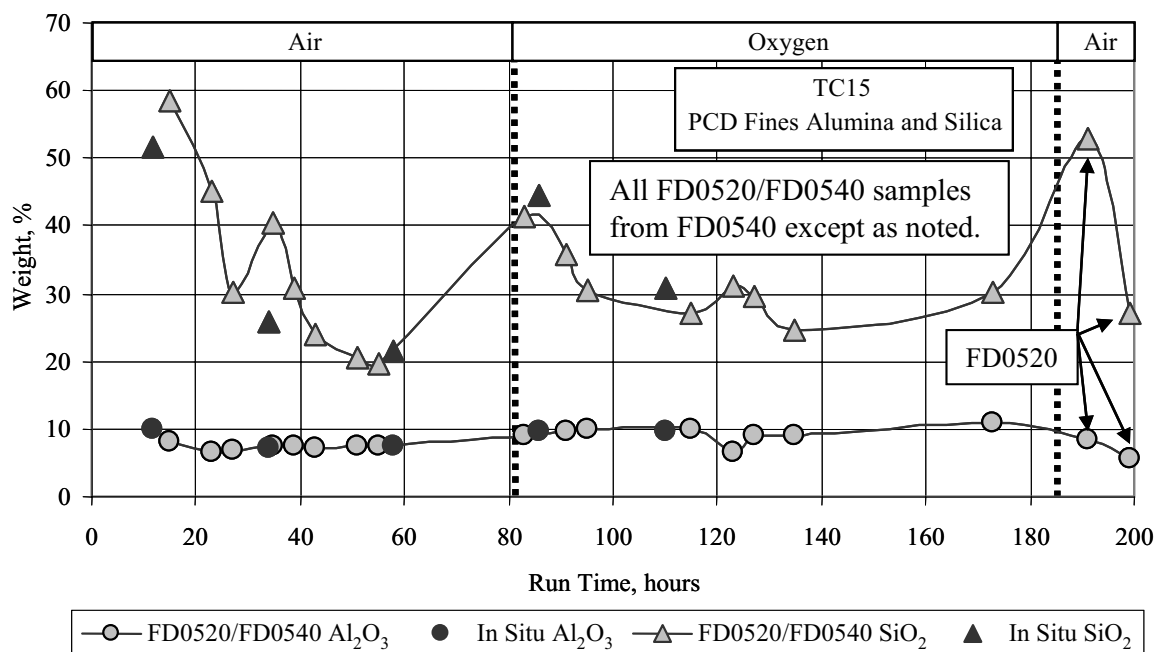


Figure 3.1-17 PCD Fines Silica and Alumina

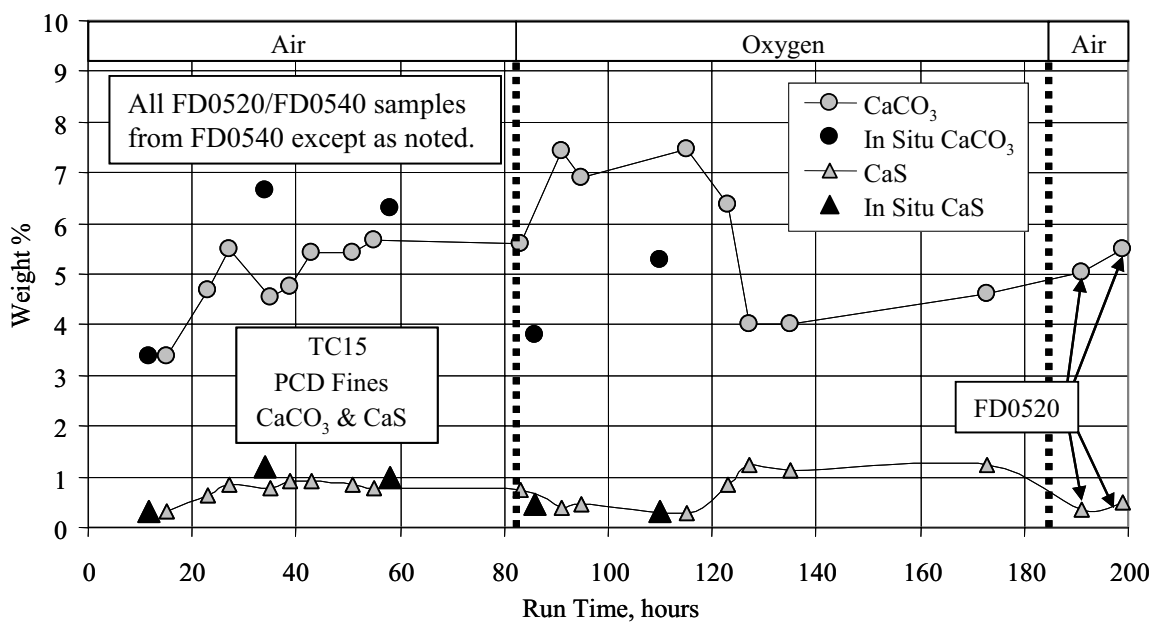


Figure 3.1-18 PCD Fines Calcium Carbonate and Calcium Sulfide

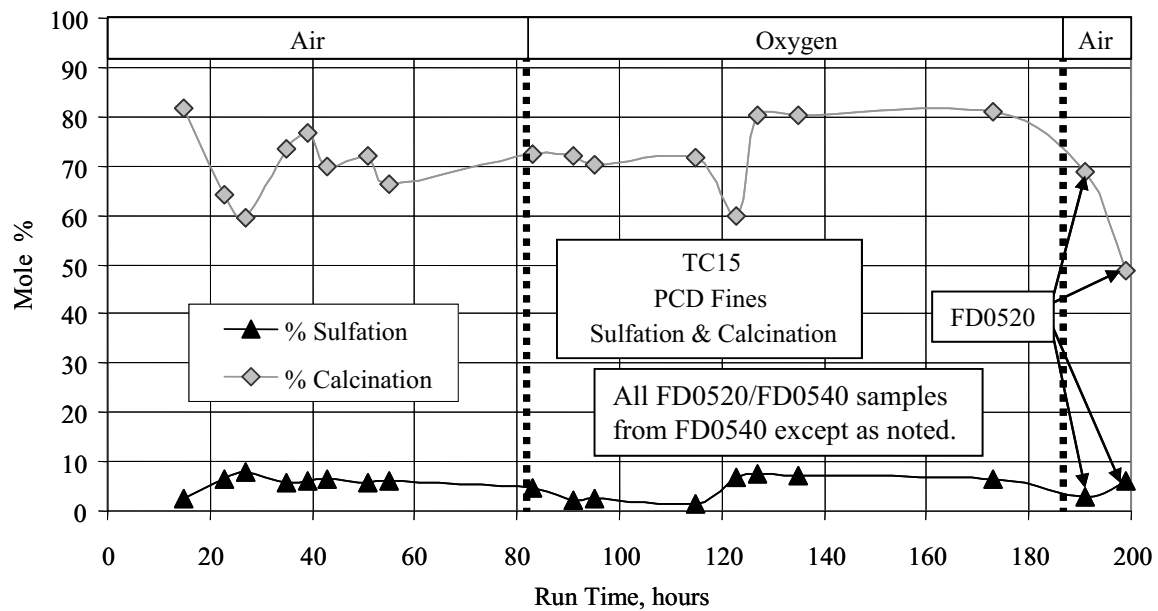


Figure 3.1-19 PCD Fines Calcination and Sulfation

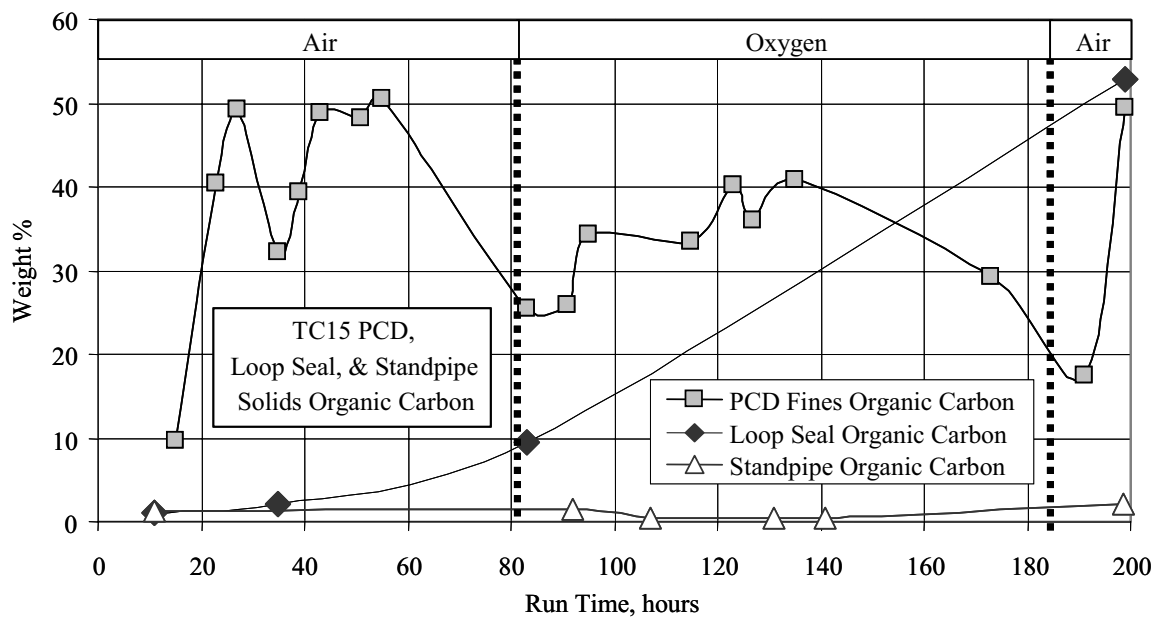


Figure 3.1-20 Standpipe and Loop Seal Solids and PCD Fines Organic Carbon Content

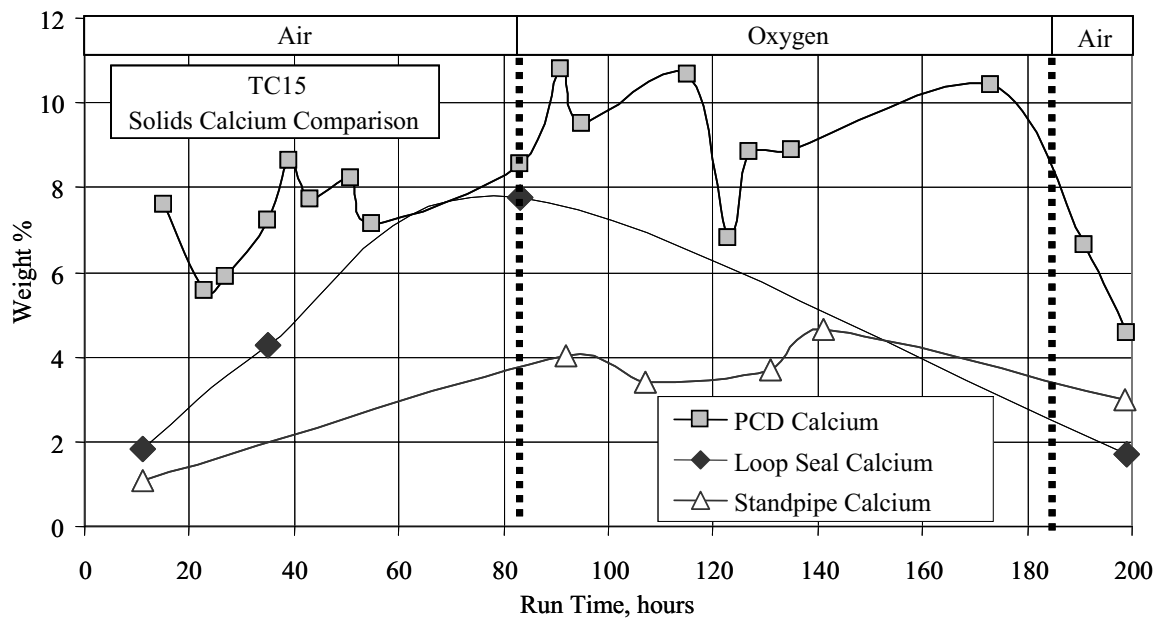


Figure 3.1-21 Standpipe and Loop Seal Solids and PCD Fines Calcium

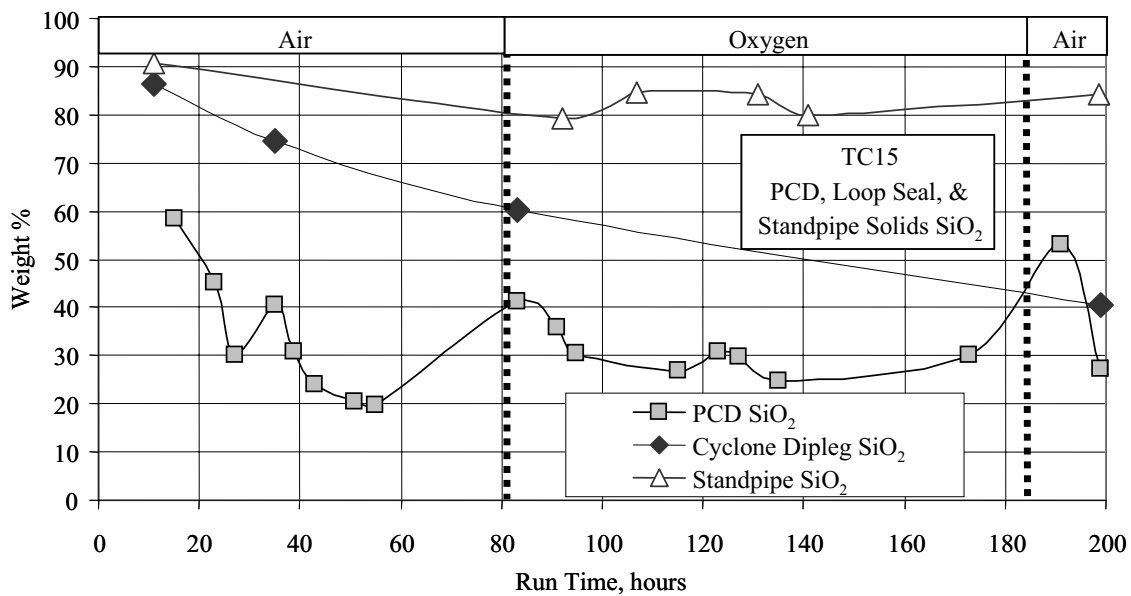


Figure 3.1-22 Standpipe and Loop Seal Solids and PCD Fines Silica

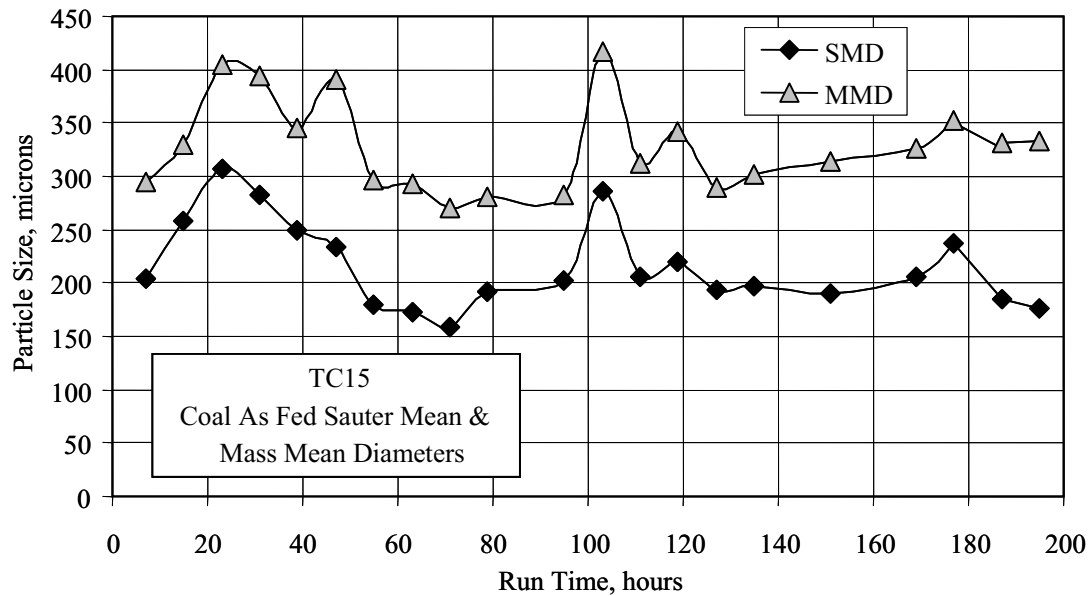


Figure 3.1-23 Coal Particle Size

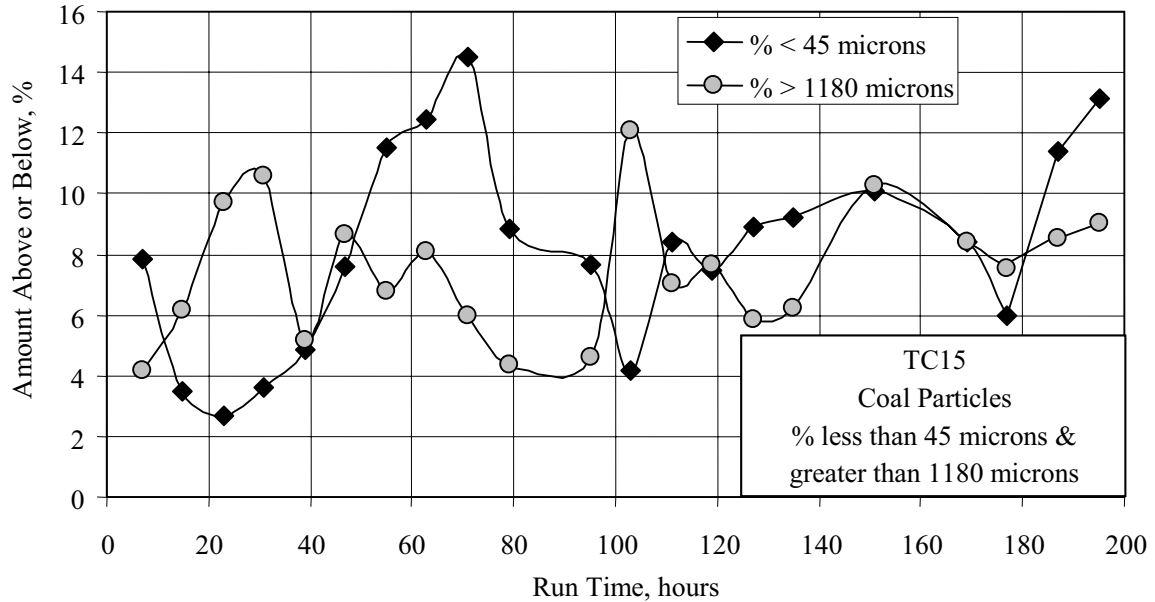


Figure 3.1-24 Percent Coal Fines and Oversize

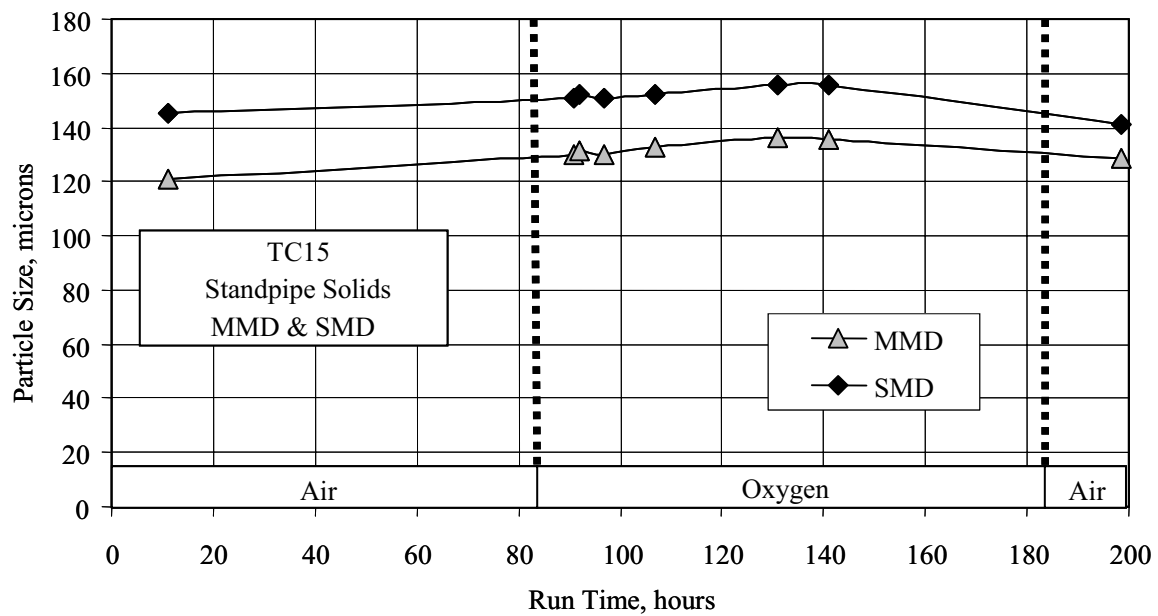


Figure 3.1-25 Standpipe Solids Particle Size

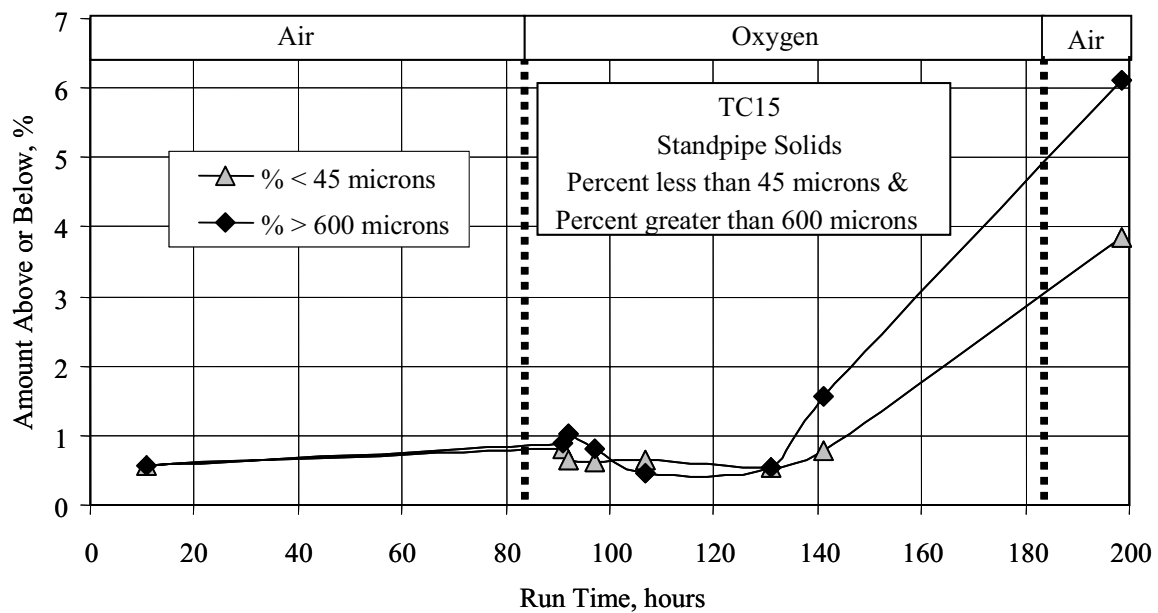


Figure 3.1-26 Standpipe Solids Fine and Coarse Particles

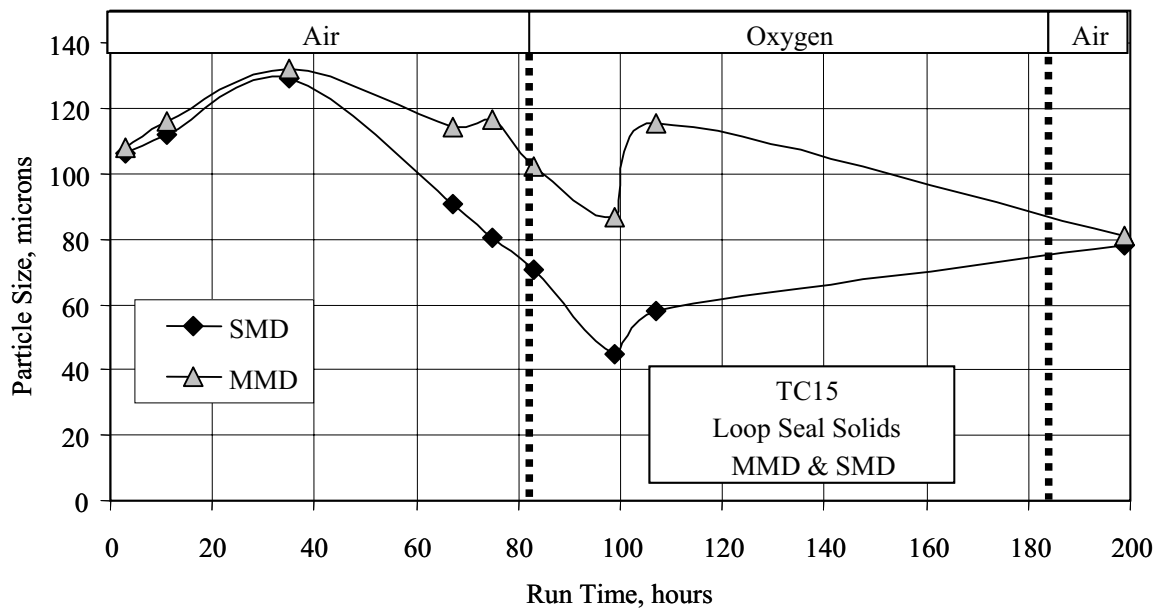


Figure 3.1-27 Loop Seal Solids Particle Sizes

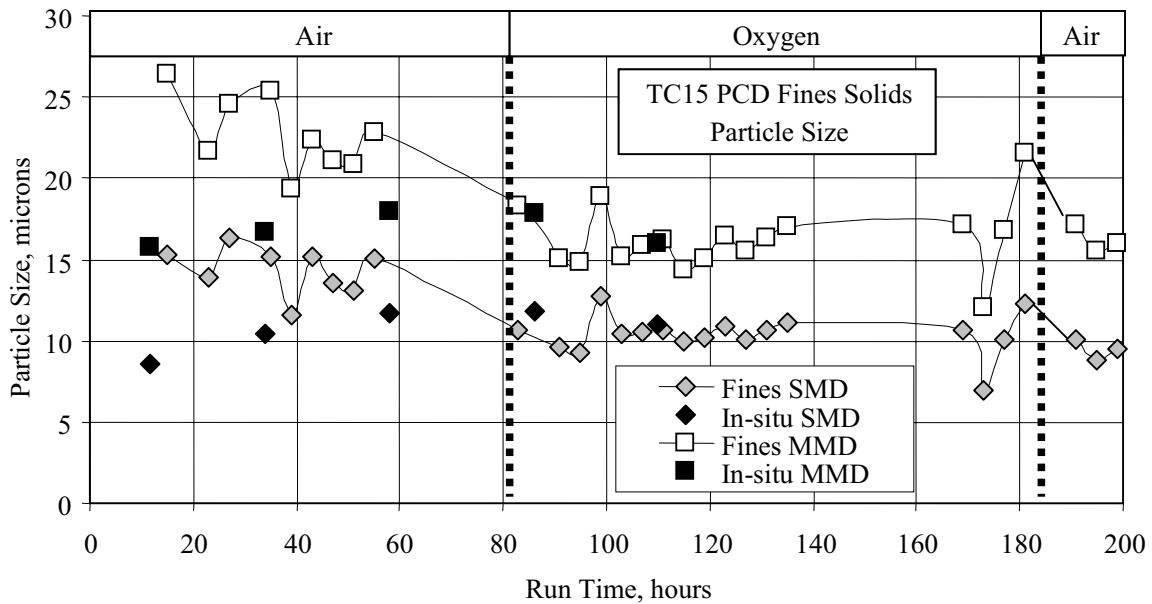


Figure 3.1-28 PCD Fines Particle Sizes

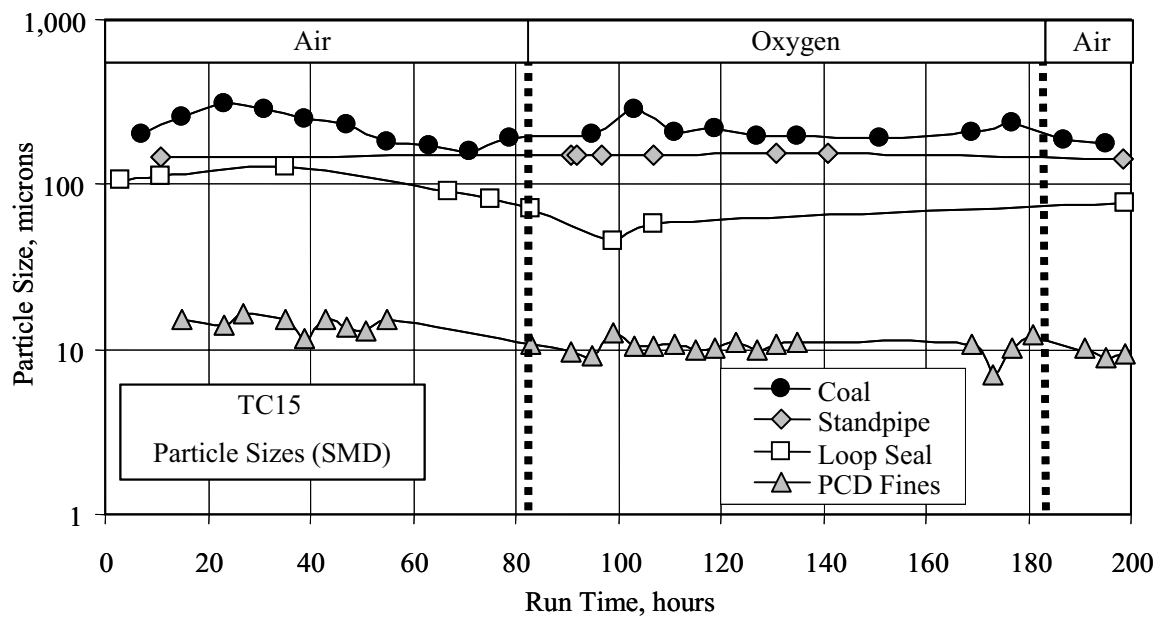


Figure 3.1-29 Particle Size Distribution

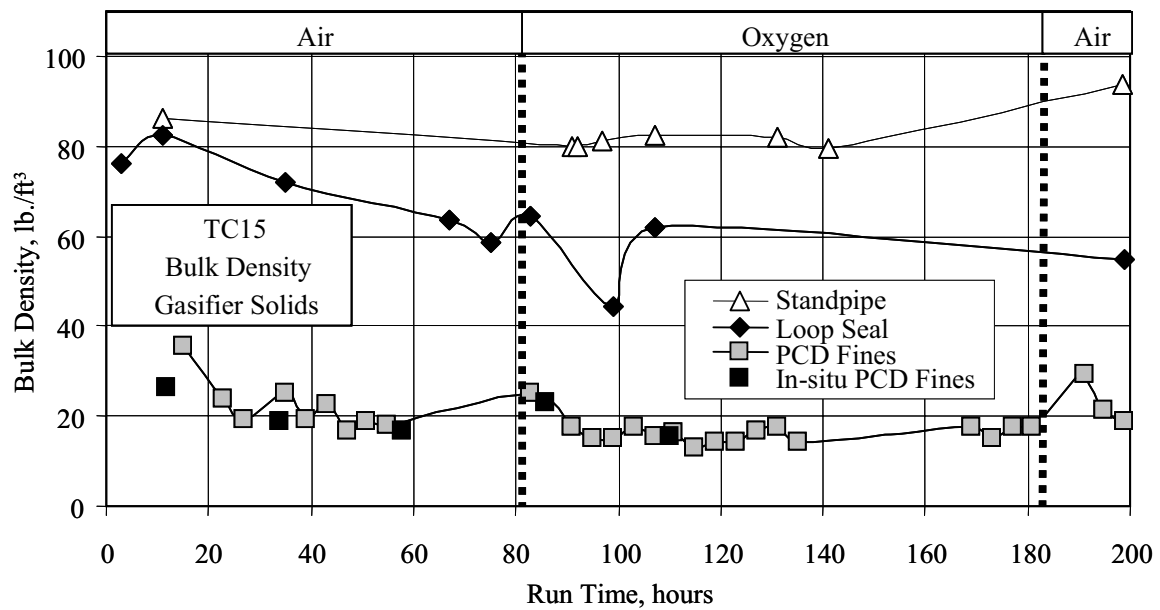


Figure 3.1-30 Standpipe and Loop Seal Solids and PCD Fines Bulk Density

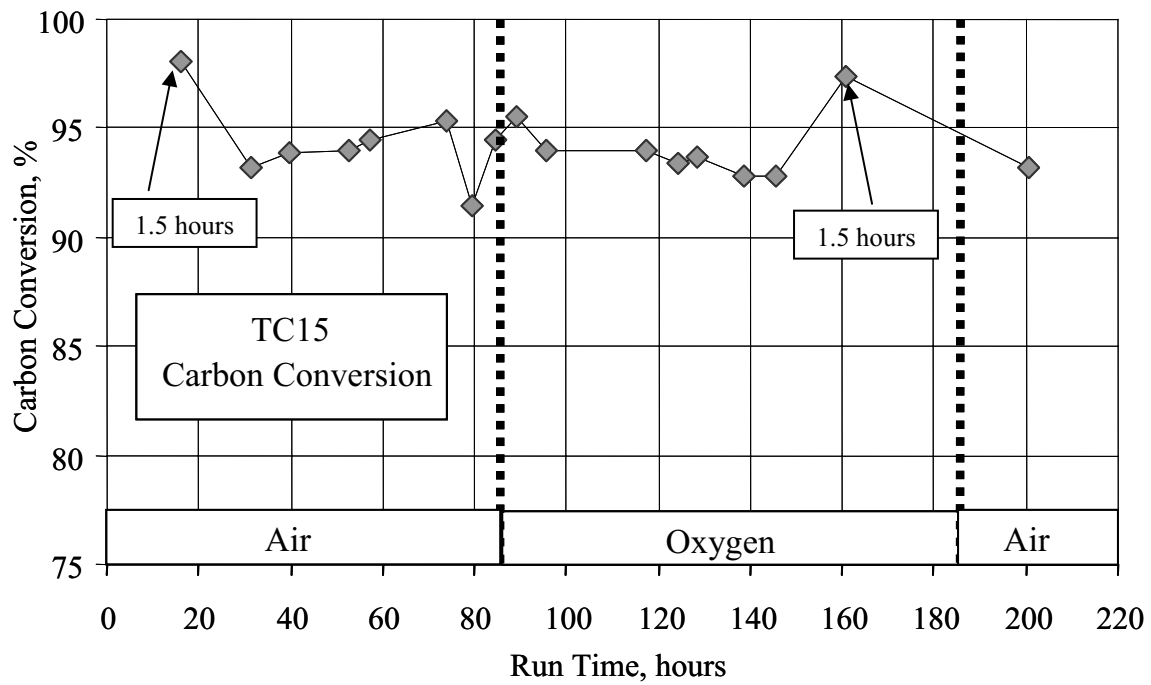


Figure 3.1-31 Carbon Conversion

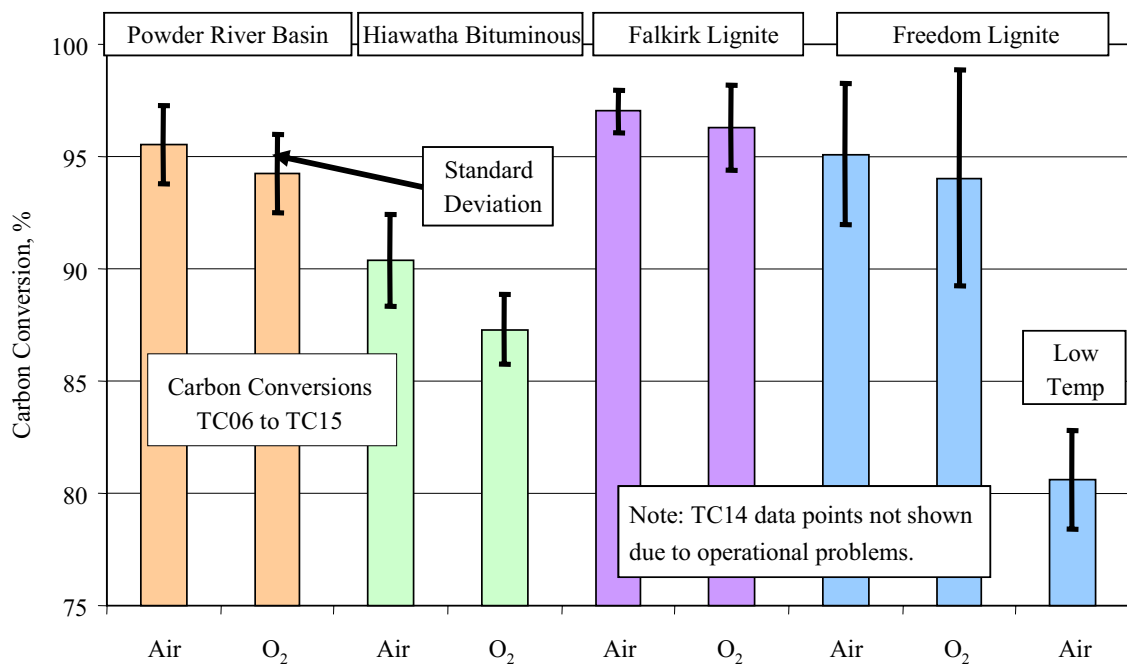


Figure 3.1-32 Carbon Conversion of Four Coals

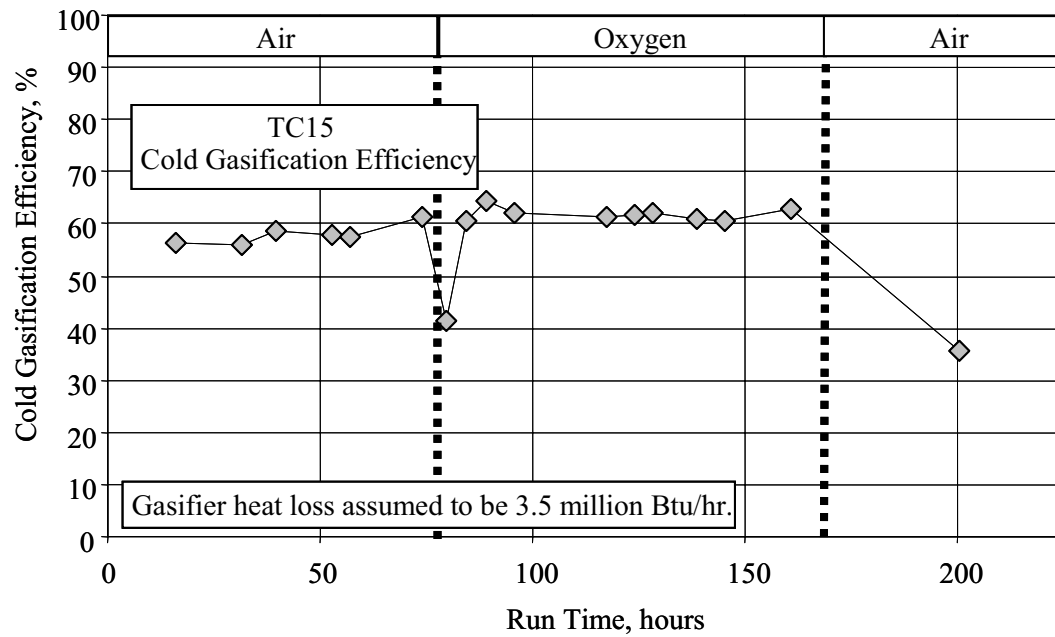


Figure 3.1-33 Cold Gasification Efficiency

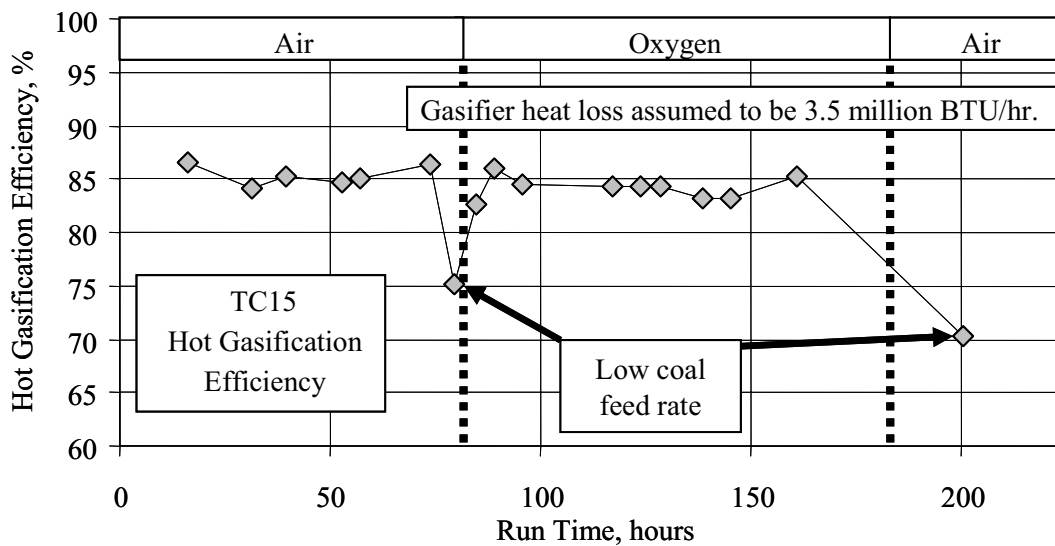


Figure 3.1-34 Hot Gasification Efficiency

3.2 PARTICULATE CONTROL DEVICE

3.2.1 Overview

Overall, operations for the PCD were quite stable throughout TC15. Despite problems with reactor circulation, no major upsets were seen in the PCD. There were no filter element failures and no char bridging. During most of the run, the baseline pressure drop was about 60 to 120 inH₂O. It was slightly higher during the times the developmental coal feeder was operating. Generally, the baseline pressure drop was slightly higher than seen in previous runs. During steady state operations, the inlet temperature was about 700 to 950°F and the face velocity was maintained at 3 to 5 ft/min.

As in TC14, a reduced number of filter elements were installed to evaluate a higher face velocity in the PCD. For this test run, 73 iron-aluminide filter elements were tested. Several of these filter elements have been exposed to syngas for approximately 4,400 hours. Evaluation of the seven Siemens Westinghouse inverted filter element assemblies continued during this test run. Failsafe devices installed included 23 Pall fuses, 3 CeraMem ceramics, and 46 PSDF-designed failsafes. In addition to these failsafes, six Siemens Westinghouse metal fiber failsafes were installed above blanks for material testing.

Outlet loading samples were collected throughout the test run. A small amount of particle penetration of approximately 0.25 ppmw was indicated during the first 2 days. Despite the initial small amount of particle penetration, outlet mass concentrations dropped below 0.1 ppmw, the lower limit of detection, for the remainder of the run. Failsafe testing with hot gasification ash injection was performed on the CeraMem ceramic failsafe. The first outlet sample collected during the failsafe testing showed approximately 0.5 ppmw, and a second outlet sample taken 4 hours later showed approximately 0.17 ppmw.

Optimization testing was performed on the back-pulse parameters to determine the effect of back-pulse pressure on the baseline pressure drop and filter element cleaning. The back-pulse pressures were varied from 150 to 400 psi above system pressure. An instrumented filter element was used to determine the actual pressure intensities across the filter media and dustcake. The tests showed no significant increase in baseline pressure drop and no evidence of inadequate filter element cleaning.

TC15 provided an opportunity to study the dust characteristics and PCD performance while gasifying PRB coal in both air- and oxygen-blown modes. It also provided an opportunity to examine the effects of the cyclone repairs on the concentration and characteristics of particles being carried over to the PCD.

3.2.2 Particle Mass Concentrations

In situ particulate sampling was performed at the inlet and outlet of the PCD to quantify the particulate collection efficiency and to relate PCD performance to the characteristics of the dust entering the PCD. The samples were collected using the SRI in situ batch sampling system described in previous reports. Measurements were made in each of the two operating modes.

Particle mass concentrations and mass rates measured at the PCD inlet are given in [Table 3.2-1](#). The mass rates are also plotted as a function of the coal-feed rate in [Figure 3.2-1](#). In this figure, the solid line is a linear regression to the TC15 data, and the dashed lines are regressions to the TC12 (no limestone) and TC14 data points. As shown, the mass rates measured in TC15 returned to normal levels after being elevated in TC14. This result confirms that the cyclone repairs that were done between TC14 and TC15 were effective in restoring the performance of the cyclone. In [Figure 3.2-1](#), the two highest TC15 data points were obtained during oxygen-blown operation. In the past, a difference between air and oxygen operation had not been seen. However, poor circulation in the reactor during oxygen-blown operation may have affected the results. The small number of data points also suggests caution in drawing any conclusions from these data.

Particle mass concentrations measured at the PCD outlet are also shown in [Table 3.2-1](#) and compared to other test programs in [Figure 3.2-2](#). As in several recent test programs, a slightly elevated particulate loading was measured on the first 2 days after startup with the concentration falling below the lower measurement limit (0.10 ppmw) on the third day. Although the collection of tar may have somewhat increased the concentration measured in Run Number 1, microscopic examination indicated a high number of dust particles on the filter. As suggested in the TC14 report, particle leakage that occurs at the startup of a test run may be the result of the loss of fastener torque on the filters or failsafes, allowing leakage around or through the gaskets. Lab tests have indicated that fiber gaskets shrink when heated to the PCD operating temperature due to burn off of the binder material and breakdown of the filler fibers in the gaskets. Presumably, the leakage paths are small enough that they eventually plug up with dust and stop leaking after a few days.

Outlet particle mass concentrations measured during the CeraMem ceramic failsafe test are displayed in Run Numbers 5 and 6 of [Table 3.2-1](#). The first run was started after 15 minutes of hot gasification ash injection and lasted for 15 minutes. This run was intended to measure the short-term performance of the failsafe. Unfortunately, the sample filter was contaminated with tar and a simple gravimetric measurement of particle concentration during this time was not possible. The second test started after 3 hours of injection and was 4 hours long. This test yielded a particulate loading of 0.17 ppmw. Microscopic comparison of these two filters, ignoring the tar contamination, suggested that the particle loading during the first test was two to three times higher than the second test. If this was the case, then the initial concentration leaking through the failsafe would have been around 0.5 ppmw. This was considerably lower than the short-term particle leakage measured on a CeraMem failsafe during TC14 (3.34 ppmw). In both cases, it appeared that the CeraMem failsafe was reducing the outlet particle loading with time. An unanticipated trip ended the test program before additional long-term data on the failsafe performance could be obtained. During TC14, the failsafe reduced the outlet loading to below 0.1 ppmw after 14 hours, so a similar reduction would have been expected in TC15.

3.2.3 Real-Time Particle Monitoring

The PCME particulate monitor was operational throughout the test run. As seen in previous tests, there was considerable noise in the PCME signal along with spikes that coincided with the PCD back-pulses.

Figure 3.2-3 shows the PCME output during the failsafe injection test. There was no indication from the data that the instrument detected the estimated 0.5 ppmw that was emitted from the failsafe during the first few minutes of injection. This was consistent with previous results that indicated the PCME was not capable of detecting particle concentrations below a few ppmw, particularly when the particle-size distribution contains few large particles. Microscopic examination indicated that there were few particles larger than 25 μm collected on the sample filter during the injection test. It was previously concluded in the TC14 report that the PCME was not stimulated by particles smaller than 25 μm .

Output traces from the PCME particulate monitor from the day before the failsafe injection started are shown in Figure 3.2-4. The PCME indicated a clear increase in output starting at about 8:30 a.m. The initial increase trended downward somewhat, but the output was still significantly elevated above the starting value for the remainder of the day. This increase was not correlated with any change in process conditions. Outlet Run Number 4 (Table 3.2-1) was conducted during the elevated period and detected no increase in particle emission from the PCD. It was not clear what the PCME was responding to, but it was clear that small increases in output were not necessarily indicative of PCD leaks.

3.2.4 PCD Solids Analysis

Gasification ash samples were collected in situ at the PCD inlet, from the PCD hopper, and from the PCD dustcake. The samples were thoroughly characterized to document particle-size distributions, physical properties, chemical composition, and flow resistance (drag). The PCD pressure drop and face velocity were monitored and used along with the measured particulate loading to determine the transient drag of the gasification ash. The transient drag values calculated by this method were compared to drag values measured in the laboratory.

3.2.4.1 Particle-Size Distributions

A Microtrac X-100 particle-size analyzer was used to measure the particle-size distributions of the in situ particulate samples collected at the PCD inlet and the PCD hopper samples used for laboratory drag measurements. Figure 3.2-5 compares differential mass particle-size distributions measured on the PCD inlet in situ samples for air- and oxygen-blown operation. This comparison indicated a higher mass concentration in most particle sizes because of the lower gas flow associated with oxygen-blown operation. When the same measurements were expressed on the basis of differential mass percentage in Figure 3.2-6, there were no significant differences between the particle-size distributions produced within the two operating modes.

Any significant particle-size differences between the selected PCD hopper samples and the in situ samples would indicate that the hopper samples used for the lab drag measurements were not representative and would give higher drag results than observed for the actual PCD. The differential mass percentage distributions for the in situ and hopper samples used for lab drag measurements are compared in Figure 3.2-7. The two sets of samples showed good agreement and there was no indication that the hopper samples were not appropriate for flow resistance measurements.

3.2.4.2 Cake Thickness and Areal Loading

A clean shutdown was performed on the PCD in which it was back-pulsed for 24 hours after coal feed was terminated. This procedure left no transient cake and a very thin residual cake on the filter elements. The cake on the top plenum was generally thicker than residual cakes from previous runs. It was easily brushed off the filter element surface and appeared to be dry and not at all sticky.

[Table 3.2-2](#) summarizes the thickness and areal loading measurements made on the residual cake and compares the measurements to those made on residual cakes from previous runs. As indicated in the table, the thickness and areal loading measurements were made on three elements from the top plenum and two elements from the bottom plenum. Although the top plenum elements had a thicker cake, the cake areal loading on the top plenum elements was actually lower than it was on the bottom plenum elements. A problem was suspected with the measurements made on the bottom plenum elements because of the unusually low porosity values calculated from the thicknesses and areal loadings. Porosity values as low as 70 percent were quite unusual for gasification ash. The measurements made on the top plenum elements were more consistent with previous measurements. Therefore, only the top plenum measurements were included in the average values for TC15. All previous test runs have produced a residual cake thickness consistently on the order of 0.01 inch. However, the TC15 residual cake was about twice as thick as residual cakes seen previously. The areal loading was also about twice the areal loading that is typically produced with PRB coal. Although these measurements differ from those made in the past, they are believed to be correct since the calculated porosity values are within the expected range.

It has previously been determined that filter elements equipped with Pall fuses produce thinner residual cakes and have the lowest flow resistance while filter elements equipped with PSDF failsafes produce thicker residual cakes and have the most flow resistance. This trend was also seen with the filter elements from this test run. Results from the past three runs suggest that the thicker cakes were related to the higher flow resistance of the PSDF failsafes, which leads to less effective back-pulse cleaning.

After the TC15 cake measurements were completed, a filter element from the top plenum was removed with the residual cake intact and examined under the optical microscope. [Figure 3.2-8](#) shows a photomicrograph of the TC15 cake along with a photomicrograph of the TC14 cake taken at the same magnification. The TC14 cake was also photographed on a filter element that was removed with the cake intact. As shown in the photographs, the TC15 cake appears to have much larger pores than the TC14 cake. The reason for this difference is unknown, but it could be related to differences in tar content. Evidence of tar deposition was noted when the PCME probe was removed after TC15 and it was covered with tar as shown in [Figure 3.2-9](#). In GCT4, large pores and fissures similar to those observed in the TC15 cake were observed in a residual cake that was glued together with tar.

3.2.4.3 Physical Properties and Chemical Compositions

In situ Samples

Tables 3.2-3 and 3.2-4 give the physical properties and chemical compositions of the in situ samples collected at the PCD inlet and composite PCD hopper samples used for laboratory drag measurements. As indicated in the tables, the first three in situ samples and the first two composite hopper samples were collected during air-blown gasification. The last two in situ samples and the last composite hopper sample were collected during oxygen-blown operation. Table 3.2-5 compares the characteristics of the in situ samples collected during air- and oxygen-blown gasification. For every one of the physical properties and chemical components shown, there is overlap in the ranges obtained during air- and oxygen-blown conditions. This overlap confirms the conclusion from previous tests that the choice of oxidant (air or oxygen) has no significant effect on the physical properties or chemical composition of the gasification ash.

Table 3.2-6 compares the average characteristics of the TC15 in situ samples to previous in situ samples produced from PRB coal without limestone addition. The characteristics of the TC15 gasification ash appear to be in line with the gasification ash characteristics from previous tests, with the exception of the TC14 gasification ash. Compared to the other PRB gasification ashes, the TC14 gasification ash had a higher bulk density, lower bulk porosity, lower surface area, and larger mean particle size. The TC14 gasification ash also contained a relatively high concentration of inerts and a relatively low concentration of calcium components. The unusual properties of the TC14 gasification ash were related to the degraded cyclone performance during that run. The more representative particle size and other characteristics of the TC15 gasification ash suggest that the cyclone repairs that were done between TC14 and TC15 were successful.

Figure 3.2-10 shows the specific-surface area of the TC15 in situ samples plotted as a function of the carbon content, along with samples from previous PRB test runs. As discussed in the TC14 report, some of the TC14 in situ samples are out of line with the previously established trend of surface area versus carbon content as a result of the degraded cyclone performance in TC14.

Composite Hopper Samples

The composite hopper sample AB14866 was clearly out of line with the other composite hopper and in situ samples. This sample had an unusually high bulk density, low porosity, low surface area, and high mass-median diameter. It also had unusually low carbon content and an unusually high amount of inerts (ash/sand). The ash minerals analysis showed that the silica content of this sample was double the silica content of the other samples, suggesting that it was biased by carryover of bed material. Because of these factors, composite hopper sample AB14866 was not included in calculating the average values in the Tables 3.2-3 and 3.2-4. The other two composite hopper samples were fairly similar to the in situ samples.

Dustcake Samples

Due to a very thin residual dustcake, separate dustcake samples from individual filter elements could not be collected. It was necessary to limit sample collection to a single bulk sample from

the top plenum and a single bulk sample from the bottom plenum. [Tables 3.2-7](#) and [3.2-8](#) give the physical properties and chemical compositions of the two bulk dustcake samples taken from the top and bottom plenums. [Table 3.2-9](#) compares these samples to the TC15 in situ particulate samples.

As seen in previous tests, the residual cake appears to be enriched in fine particles (6 to 7 μm MMD in the cake versus 16 to 18 μm MMD in the in situ samples). The mechanism of the fine-particle enrichment has been discussed in previous reports and is further supported by the TC15 results. Other than the difference in particle size, the properties of the TC15 residual cake are not dramatically different from the properties of the incoming gasification ash. The dustcake samples have a slightly lower concentration of calcium components, but this difference is not considered to be significant. As discussed in previous reports, the surface area of the residual cake is usually lower than the surface area of the in situ samples, but this was not the case in TC15. In previous tests, the lower surface area of the cake has been attributed in part to closure of pores associated with the additional reaction with H_2S and formation of CaS or with tar filling pores. In TC15, this effect was apparently less significant than it was in the previous tests.

[Table 3.2-10](#) compares the average properties of the TC15 residual dustcake with those from other test campaigns that used PRB coal. The properties of the TC06 and TC12 dustcakes were apparently affected by the addition of limestone at the end of the run, and the properties of the TC14 dustcake may have been affected by the degraded cyclone performance. Therefore, the TC10 dustcake probably provides the best comparison with the TC15 dustcake. The properties and compositions of these two cakes are very similar, except for the unusually large surface area of the TC15 cake. The TC06 cake had an even higher surface area. The TC06 in situ samples also had very high surface areas, and there was some char bridging at the end of TC06. If the dustcake samples were contaminated with bridged material that was similar to the in situ samples, the bridging contamination could account for the unusually high surface area of the TC06 cake. However, this cannot explain the high surface area of the TC15 cake, as there was no bridging during TC15.

3.2.4.4 Measurements of Dustcake Flow Resistance

Drag measurements were made on two of the composite hopper samples (AB14867 and AB14868) shown in [Tables 3.2-3](#) and [3.2-4](#). Since previous measurements have shown that the carbon content of the gasification ash has an influence on drag, composite hopper samples were prepared to represent two distinct carbon levels, nominally 49- and 34-weight percent. The sample containing 49-weight percent carbon was generated during air-blown operation; while the sample containing 34-weight percent carbon was generated during oxygen-blown gasification. As shown previously, the type of oxidant (air versus oxygen) does not have a direct effect on particulate characteristics. However, the particulate characteristics, especially drag, are affected by the carbon content. Analysis of the filter catches from the laboratory drag apparatus showed LOI values that were consistent with these carbon levels.

The drag results as a function of particle size for the two composite hopper samples are shown in [Figure 3.2-11](#). The solid lines on the figure are the linear regression fits of the drag versus particle-size data. The higher-carbon sample had higher drag as expected, but the difference was not very great and was not statistically significant. This small difference was consistent with

previous data that showed large changes in drag as carbon content went from 2 to 10 percent, but much smaller changes at higher carbon contents. The dashed line is the average correlation predicted by a multiple regression model that takes into account both the particle size and the carbon content of the samples. This dashed line is probably the best representation of the TC15 drag data.

The dotted lines on [Figure 3.2-11](#) show the range of previous drag data with PRB coal. The drag data from TC15 were quite high and were at the upper end of the range of previous data with PRB coal. This data cast doubt on the conclusion that was made in previous tests that the addition of the new LMZ in TC07 produced a permanent downward shift in drag. The true reason for the shift is unknown, but it could be related to subtle changes in the coal properties or gasifier operating conditions.

3.2.5 Analysis of PCD Pressure Drop

Transient Drag

During each in situ sampling run, the PCD transient drag was calculated using the measured pressure drop, gas flow, and particle concentration. See [Table 3.2-11](#). The calculated transient drag at PCD conditions is listed under the column heading “PCD.” The corresponding normalized value of transient drag at room temperature is listed under the heading “PCD@RT.” This value can be compared directly with the lab-measured drag values.

As noted in the previous section, lab drag measurements are affected by the carbon content of the dust. The PCD transient drag values (corrected to room temperature) are plotted as a function of non-carbonate carbon (NCC) content in [Figure 3.2-12](#), along with data from previous PRB runs. There is considerable scatter in the data due to variations in equipment configuration, process conditions, coal composition, and limestone addition. Nevertheless, the data shows a definite trend toward increasing drag with increasing carbon content. Thus, both the analysis of PCD pressure drop and the lab drag measurements show that drag is affected by the carbon content of the gasification ash.

The NCC values and particle size (MMD) used to calculate the lab drag for each PCD test condition are shown in [Table 3.2-11](#) along with the calculated drag from the multiple regressions labeled “RAPTOR.” This value of RAPTOR drag is based on the multiple regression equation derived from the drag data, MMD data, and NCC data. As shown in [Figure 3.2-13](#), the multiple regression using the actual RAPTOR filter catch carbon content yields drag values that track the normalized PCD transient drag at room temperature (PCD@RT).

[Figure 3.2-13](#) compares the values of PCD@RT to the corresponding individual values of RAPTOR drag calculated from the multiple regression model described above. Although there is again considerable scatter in the data, the data points are generally scattered around the perfect agreement line. Overall, the cumulative drag data based on RAPTOR and on actual PCD performance continue to show fairly good agreement. Both the lab measurements and PCD ΔP continue to show that there is a strong variation of drag with carbon content of the gasification ash.

Since the lab drag measurements are based on only a few bulk samples collected during “typical” operating conditions, they cannot accurately represent all of the individual test conditions encountered during a test campaign. It is this comparison of a few lab samples to a wide range of actual PCD data that causes much of the scatter in [Figure 3.2-13](#). A more accurate comparison might be to compare the average lab measurements with the average PCD data from a test program. Using average values would also simplify the presentation of the data; using all the individual data points produces a graph that is becoming quite cluttered. Accordingly, the average lab and PCD drag values for each test program are calculated and presented in [Table 3.2-12](#) and [Figure 3.2-14](#). There is still some scatter in the data, but they are almost symmetrically scattered around the perfect agreement line. To illustrate the accuracy of the lab simulation, a linear regression and 95-percent confidence intervals to the mean are also plotted on [Figure 3.2-14](#). Although the regression to the data and the perfect agreement line do not exactly coincide, the perfect agreement line does fall inside the 95-percent confidence interval, indicating very good agreement between the data sets.

Baseline Pressure Drop

During most of the run, the baseline pressure drop was about 60 to 120 inches of water. The reduction in filter surface area resulted in a slightly higher baseline pressure drop compared to previous test runs with a higher filter surface area. The baseline pressure drop and coal-feed rate during the test run are plotted in [Figure 3.2-15](#). Overall, the baseline pressure drop corresponded to the coal-feed rate. As the coal-feed rate was increased, the baseline pressure drop increased. The most stable pressure drop was observed during the time in which the back-pulse pressure was lowered to a 150 psi differential during back-pulse optimization testing. It remained steady at about 80 inches of water until circulation problems in the gasifier were encountered. As shown in the graph, the baseline pressure drop during the last few days of the test run was unsteady. The erratic pressure drop was contributed to circulation problems in the gasifier, high solids carryover to the PCD, and the brief period of time coal feed was terminated and the startup burner was used.

Table 3.2-1

PCD Inlet and Outlet Particulate Measurements for TC15

Test Date	PCD Inlet					PCD Outlet				
	Run No.	Start Time	End Time	Particle Loading,		Run No.	Start Time	End Time	H ₂ O Vapor, vol %	Particle Loading, ppmw
				ppmw	lb/hr					
Air Blown - PRB										
4/20/04	1	10:30	10:45	9700	173	1	10:15	14:15	10.5	< 0.25 ⁽¹⁾
4/21/04	2	9:20	9:35	18500	366	2	8:45	12:45	(2)	0.12
4/22/04	3	9:00	9:15	16900	236	3	8:45	13:15	10.1	< 0.10
Oxygen Blown - PRB										
4/23/04	4	12:25	12:40	29700	367	4	10:15	14:15	22.5	< 0.10
4/24/04	5	12:30	12:45	23700	365	5	9:37	9:52	25.8	(3)
4/24/04	--	--	--	--	--	6	12:22	16:22	30.6	0.17 ⁽⁴⁾
Notes: 1. Both fine particulate and light yellow tar. Upper limit value. 2. Water sample invalid. 3. Ceramem FS injection 0:15-0:30. More tar on filter than particles. No result. 4. Ceramem FS injection 3:00-7:00.										

Table 3.2-2

Residual Cake Measurements From TC15 and Previous Runs

Element No.	Element Type	Failsafe Type	Thickness, in.	Areal Loading, lb/ft ²	Porosity, %
T-2	FEAL	PSDF	0.023	0.0236	91.0
T-3	FEAL	PSDF	0.031	0.0227	93.6
T-9	FEAL	PSDF	0.021	0.0240	90.0
B-5	FEAL	Pall Fuse	0.013	0.0450	69.5
B-7	FEAL	CeraMem Ceramic	0.016	0.0549	70.6
TC15 Average (Top Plenum Only)			0.025	0.0234	91.5
TC14 Average			0.014	0.0091	94.3
TC13 Average			0.011	0.0130	87.8
TC12 Average			0.011	N.M.*	N.M.*
TC11 Average			0.013	0.0230	84.5
TC10 Average			0.010	N.M.*	N.M.*
TC09 Average			0.008	0.0060	93.3
TC08 Average			0.010	N.M.*	N.M.*
TC07 Average			N.M.*	N.M.*	N.M.*
TC06 Average			0.010	0.0200	82.5
* N.M. = Not Measured.					

Table 3.2-3

Physical Properties of TC15 In situ Samples and Hopper Samples Used for RAPTOR

Sample ID	Run No.	Sample Date	Bulk Density g/cc	True Density g/cc	Uncompacted Bulk Porosity %	BET Specific Surface Area m ² /g	Mass Median Diameter μm
<i>In-Situ Samples, Air-Blown</i>							
AB14854	1	4/20/04	0.42	2.50	83.2	68	15.8
AB14855	2	4/21/04	0.30	2.18	86.2	193	16.7
AB14856	3	4/22/04	0.27	1.88	85.6	186	18.0
<i>In-Situ Samples, Oxygen-Blown</i>							
AB14857	4	4/23/04	0.37	2.32	84.1	102	17.8
AB14858	5	4/24/04	0.25	2.15	88.4	181	16.0
<i>Average</i>			0.32	2.21	85.5	146	16.9
<i>Composite Hopper Samples Used for Lab Drag Measurements, Air-Blown</i>							
AB14866	---	4/20/04	1.01	2.61	61.3	21	55.1
AB14867	---	4/22/04	0.32	2.10	84.8	128	21.8
<i>Composite Hopper Samples Used for Lab Drag Measurements, Oxygen-Blown</i>							
AB14868	---	4/24/04	0.28	2.35	88.1	146	15.3
<i>Average (excludes AB14866)</i>			0.30	2.23	86.4	137	18.5

Table 3.2-4

Chemical Composition of TC15 In situ Samples and Hopper Samples Used for RAPTOR

Sample ID	Run No.	Sample Date	CaCO ₃ Wt %	CaS Wt %	CaO Wt %	Non-Carbonate Carbon Wt %	Inerts (Ash/Sand) Wt %	Loss on Ignition Wt %
<i>In-Situ Samples, Air-Blown</i>								
AB14854	1	04/20/04	3.39	0.04	11.83	14.39	70.34	13.45
AB14855	2	04/21/04	6.66	1.16	5.89	43.83	42.46	46.95
AB14856	3	04/22/04	6.30	0.87	6.19	48.15	38.49	51.46
<i>In-Situ Samples, Oxygen-Blown</i>								
AB14857	4	04/23/04	3.80	0.27	11.34	20.09	64.50	21.65
AB14858	5	04/24/04	5.27	0.22	9.21	33.97	51.33	37.56
<i>Average</i>			<i>5.08</i>	<i>0.51</i>	<i>8.89</i>	<i>32.09</i>	<i>53.43</i>	<i>34.21</i>
<i>Composite Hopper Samples Used for Lab Drag Measurements, Air-Blown</i>								
AB14866	1	04/20/04	2.09	0.00	5.17	4.52	88.22	13.45
AB14867	2	04/22/04	5.39	0.71	6.74	48.81	38.35	46.95
<i>Composite Hopper Samples Used for Lab Drag Measurements, Oxygen-Blown</i>								
AB14868	3	04/24/04	5.18	0.25	9.80	33.79	50.98	51.46
<i>Average (excludes AB14866)</i>			<i>5.28</i>	<i>0.48</i>	<i>8.27</i>	<i>41.30</i>	<i>44.66</i>	<i>49.21</i>

Table 3.2-5

Comparison of In situ Sample Physical Properties During Air- and Oxygen-Blown Gasification

Oxidant	Air	Oxygen
Bulk density, g/cc	0.27-0.42	0.25-0.37
Skeletal particle density, g/cc	1.88-2.50	2.15-2.32
Uncompacted bulk porosity, %	83.2-86.2	84.1-88.4
Specific surface area, m ² /g	68-193	102-181
Mass-median diameter, μm	15.8-18.0	16.0-17.8
Non-carbonate carbon, wt %	14.4-48.2	20.1-34.0
CaCO ₃ + CaS + CaO, wt %	13.4-15.3	14.7-15.4
Inerts (ash/sand), wt %	38.5-70.3	51.3-64.5

Table 3.2-6

Comparison of In situ Sample Physical Properties to Previous Test Runs

Physical Property	TC15	TC14	TC12	TC10	TC08
Bulk density, g/cc	0.32	0.38	0.27	0.27	0.25
Skeletal particle density, g/cc	2.21	2.29	2.34	2.25	2.37
Uncompacted bulk porosity, %	85.5	83.5	88.5	88.0	89.3
Specific surface area, m ² /g	146	131	166	146	223
Mass-median diameter, μm	16.9	21.0	16.2	12.3	18.6
Non-carbonate carbon, wt %	32.1	33.4	34.1	39.4	37.5
CaCO ₃ + CaS + CaO, wt %	14.5	9.6	14.2	11.4	14.3
Inerts (ash/sand), wt %	53.4	57.0	50.7	49.2	48.3

Table 3.2-7

Physical Properties of TC15 Residual Dustcake

Sample ID	Sample Date	Bulk Density g/cc	True Density g/cc	Uncompacted Bulk Porosity %	BET Specific Surface Area m ² /g	Mass Median Diameter μm
Top Plenum						
AB14862	5/6/04	0.29	1.95	85.1	203	5.8
Bottom Plenum						
AB14863	5/6/04	0.32	1.96	83.7	160	7.3
Average		0.31	1.96	84.4	182	6.5

Table 3.2-8

Chemical Composition of TC15 Residual Dustcake

Sample ID	Sample Date	CaCO ₃ Wt %	CaS Wt %	CaO Wt %	Non-Carbonate Carbon Wt %	Inerts (Ash/Sand) Wt %	Loss on Ignition Wt %
Top Plenum							
AB14862	5/6/04	3.95	1.30	3.62	44.51	46.61	55.09
Bottom Plenum							
AB14863	5/6/04	1.75	1.25	4.74	46.76	45.50	56.23
Average		2.85	1.28	4.18	45.63	46.06	55.66

Table 3.2-9

Comparison of Bulk Dustcake Samples to In situ Samples

Physical Property	Top Plenum Cake	Bottom Plenum Cake	In Situ Range	In Situ Average
Bulk density, g/cc	0.29	0.32	0.25-0.42	0.32
Skeletal particle density, g/cc	1.95	1.96	1.88-2.50	2.21
Uncompacted bulk porosity, %	85.1	83.7	83.2-88.4	85.5
Specific surface area, m ² /g	203	160	68-193	146
Mass-median diameter, μm	5.8	7.3	15.8-18.0	16.9
Non-carbonate carbon, wt %	44.5	46.8	14.4-48.2	32.1
CaCO ₃ + CaS + CaO, wt %	8.9	7.7	13.4-15.4	14.5
Inerts (ash/sand), wt %	46.6	45.5	38.5-70.3	53.4

Table 3.2-10

Comparison of TC15 Residual Dustcake Physical Properties to Other Test Runs

Physical Property	TC15	TC14	TC12	TC10	TC06
Limestone added at end of run	No	No	Yes	No	Yes
Bulk density, g/cc	0.31	0.28	0.29	0.23	0.25
Skeletal particle density, g/cc	1.96	2.52	2.27	2.06	2.28
Uncompacted bulk porosity, %	84.4	88.9	87.2	88.8	89.0
Specific surface area, m ² /g	182	73	82	92	257
Mass-median diameter, μm	6.5	6.7	9.6	4.5	9.3
Non-carbonate carbon, wt %	45.6	26.5	26.8	49.6	40.1
CaCO ₃ + CaS + CaO, wt %	8.3	13.6	23.0	10.0	25.7
Inerts (ash/sand), wt %	46.1	60.0	50.2	40.4	34.2

Table 3.2-11

TC15 Transient Drag Determined From PCD ΔP and From RAPTOR

Run No.	$\Delta P/\Delta t$, inwc/min	$\Delta(Al)/\Delta t$, lb/ft ² /min	FV, ft/min	MMD, μm	NCC, %	Drag, inwc/(lb/ft ²)/(ft/min)		
						PCD	PCD@RT	RAPTOR
Air-Blown - PRB								
1	0.73	0.013	4.03	15.8	14.4	54	32	81
2	3.10	0.028	4.18	16.7	43.8	110	63	77
3	3.25	0.018	4.08	18.0	48.2	178	103	72
Oxygen-Blown - PRB								
4	1.03	0.028	3.04	17.8	20.1	36	23	72
5	2.33	0.028	3.76	16.0	34.0	83	52	80
Note: RAPTOR drag data calculated from multiple regression to MMD and NCC.								

Nomenclature:

$\Delta P/\Delta t$ = rate of pressure drop rise during particulate sampling run, inwc/min.

$\Delta(AL)/\Delta t$ = rate of increase in areal loading during sampling run, lb/min/ft².

FV = average PCD face velocity during particulate sampling run, ft/min.

MMD = mass-median diameter of in situ particulate sample, μm .

NCC = non-carbonate carbon.

RT = room temperature, 77°F (25°C).

RAPTOR = resuspended ash permeability tester.

Table 3.2-12

Comparison of Average Drag Values Determined From PCD Performance and From Lab Measurements

Run	Coal	Average Transient Drag Determined from PCD Performance, inwc/(lb/ft ²)/(ft/min)	Average Drag Determined from RAPTOR Lab Measurements, inwc/(lb/ft ²)/(ft/min)	Difference, %
GCT2	PRB	29.3	20.9	-33.5
GCT3	PRB	80.2	92.7	14.5
GCT4	PRB	66.4	57	-15.2
TC06	PRB	89.4	81.2	-9.6
TC07	PRB	47.7	49.8	4.3
TC08	PRB	46.5	50	7.3
TC09	Hiawatha Bit	29.0	23.3	-21.8
TC10	PRB	44.7	57.6	25.2
TC11	Falkirk Lignite	16.1	35.9	76.2
TC12	PRB	58.0	60.8	4.7
TC13	Freedom Lignite	34.4	39.4	13.6
TC14	PRB	47.4	41.6	-13.0
TC15	PRB	54.6	76.4	33.3
<i>Average</i>		49.5	52.8	6.6

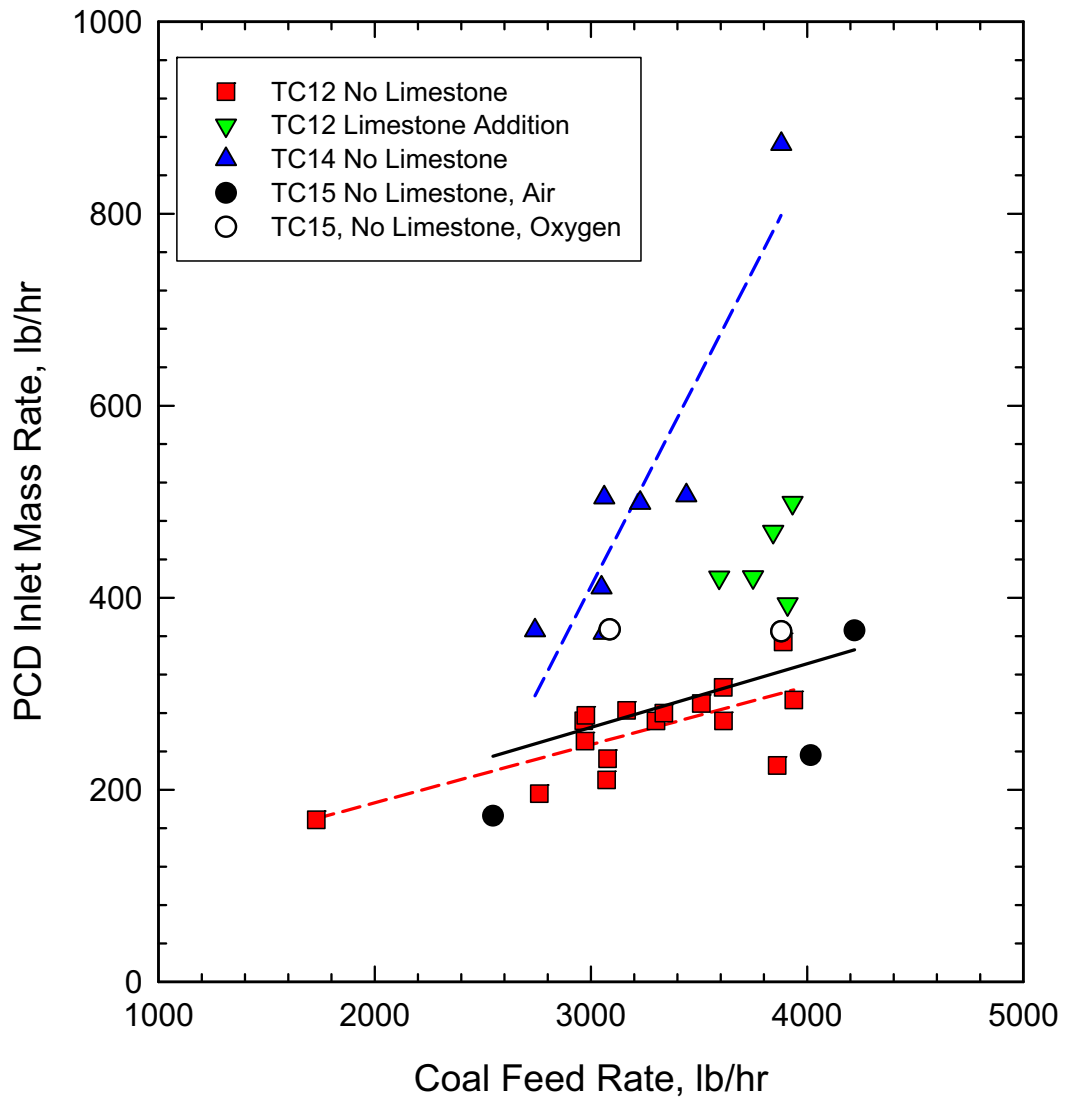


Figure 3.2-1 PCD Inlet Particle Concentration as a Function of Coal-Feed Rate

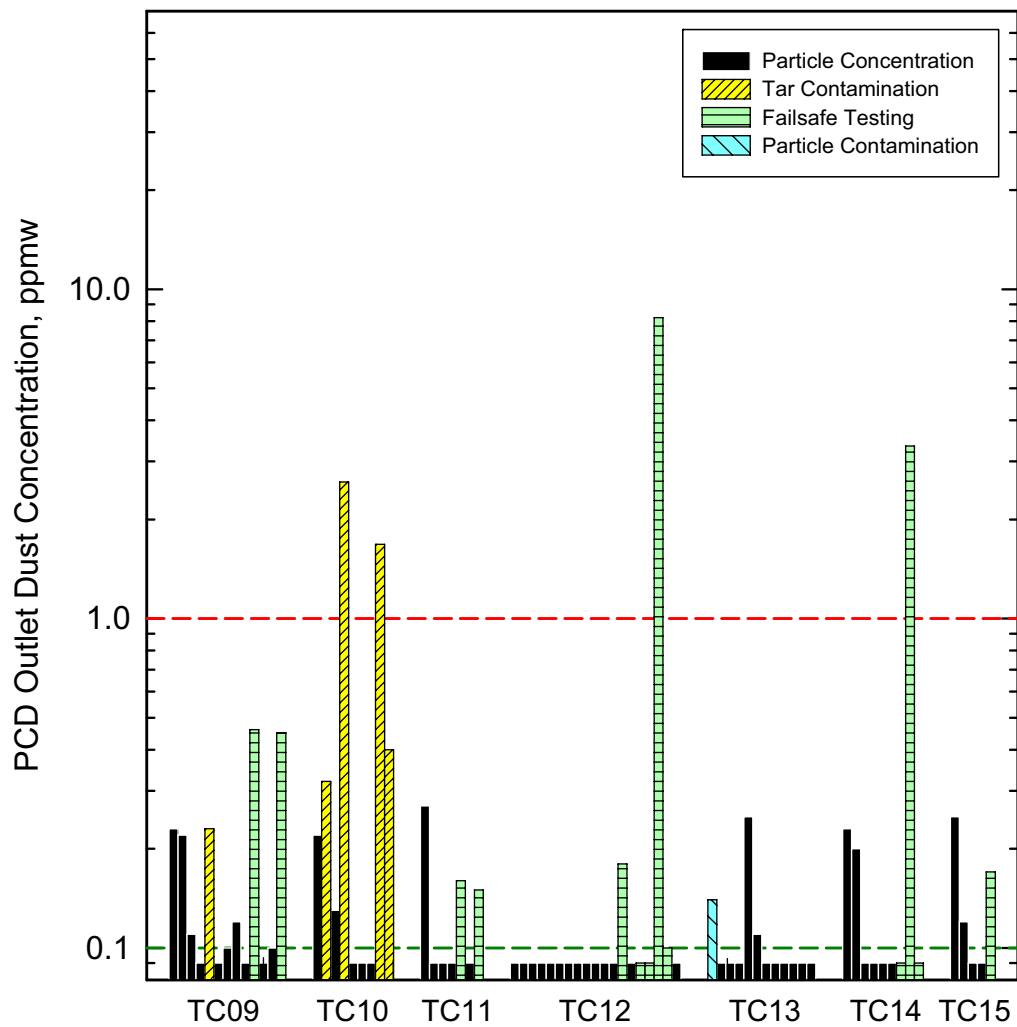


Figure 3.2-2 PCD Outlet Emissions for Recent Gasification Runs

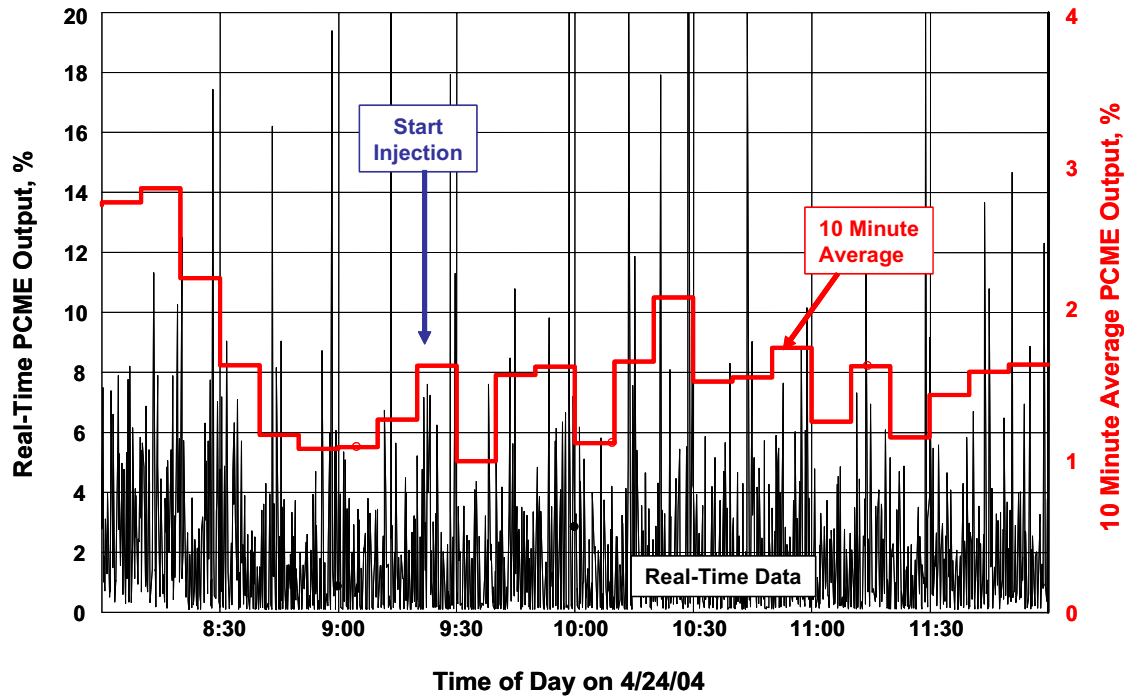


Figure 3.2-3 PCME Output During Failsafe Injection Test

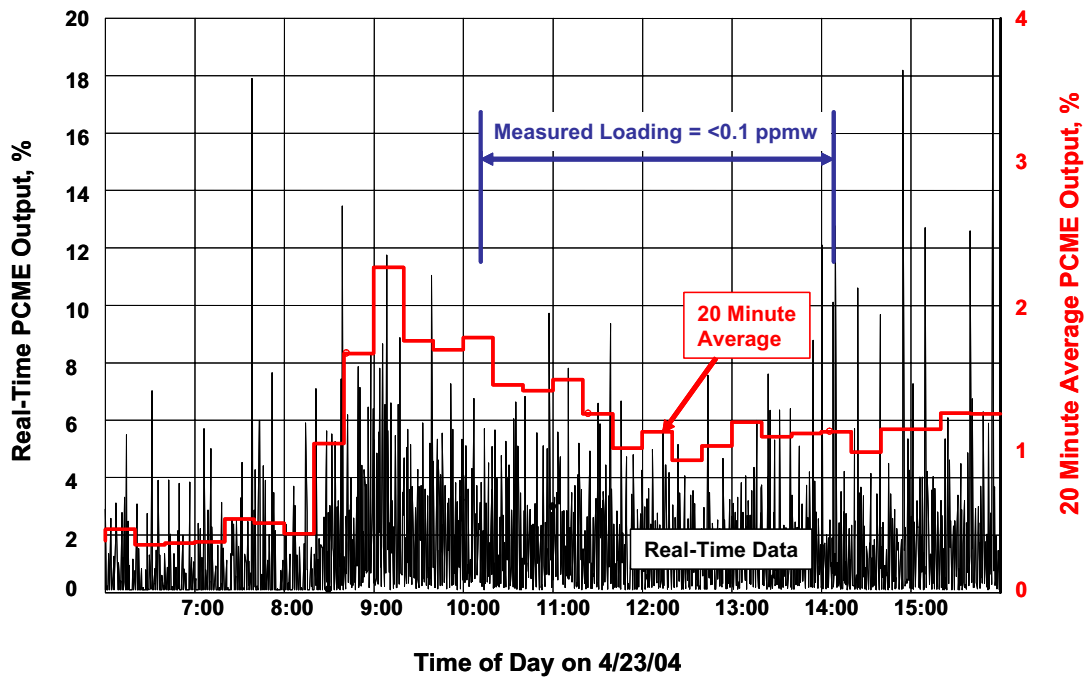


Figure 3.2-4 Elevated PCME Output With Unknown Cause

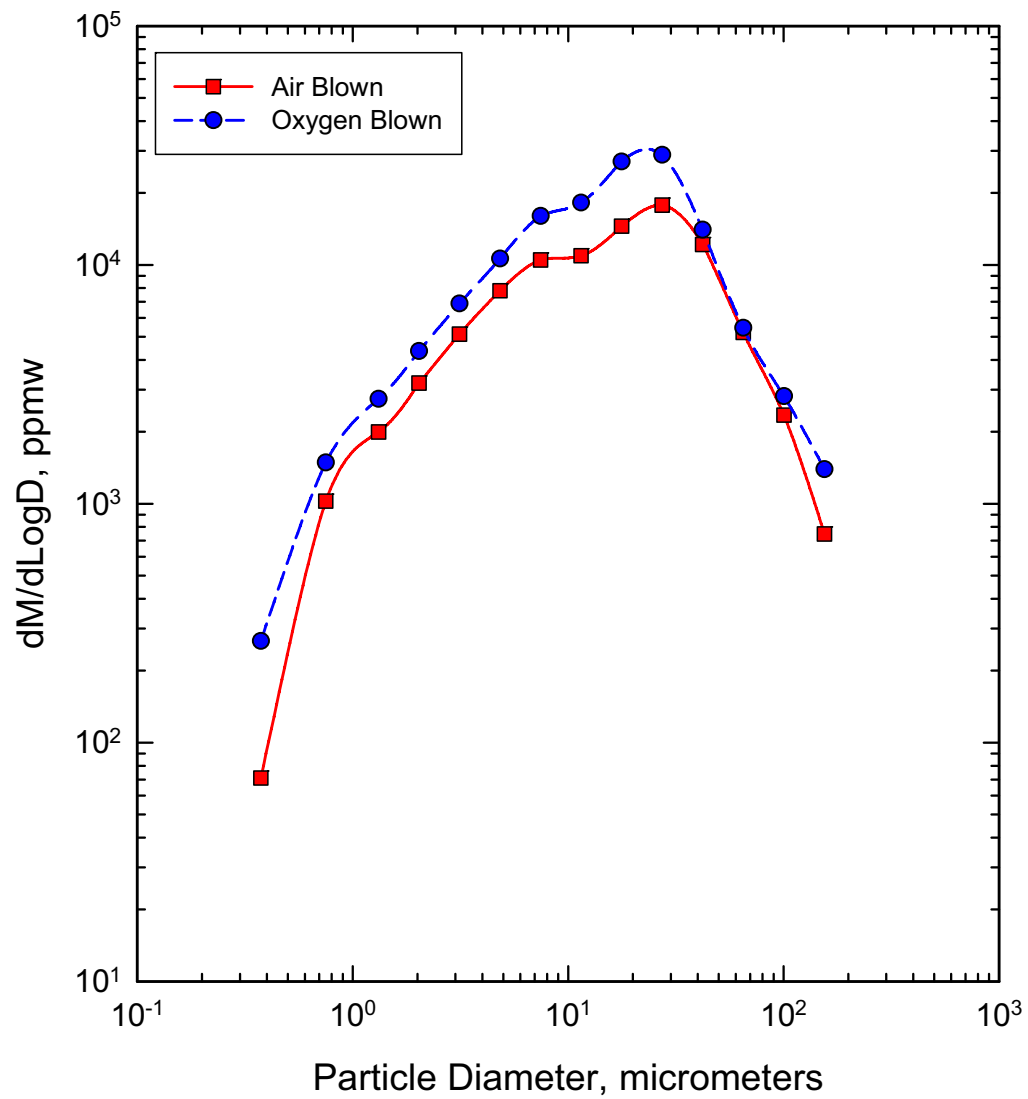


Figure 3.2-5 Comparison of Average PCD Inlet Particle-Size Distributions on Mass Basis

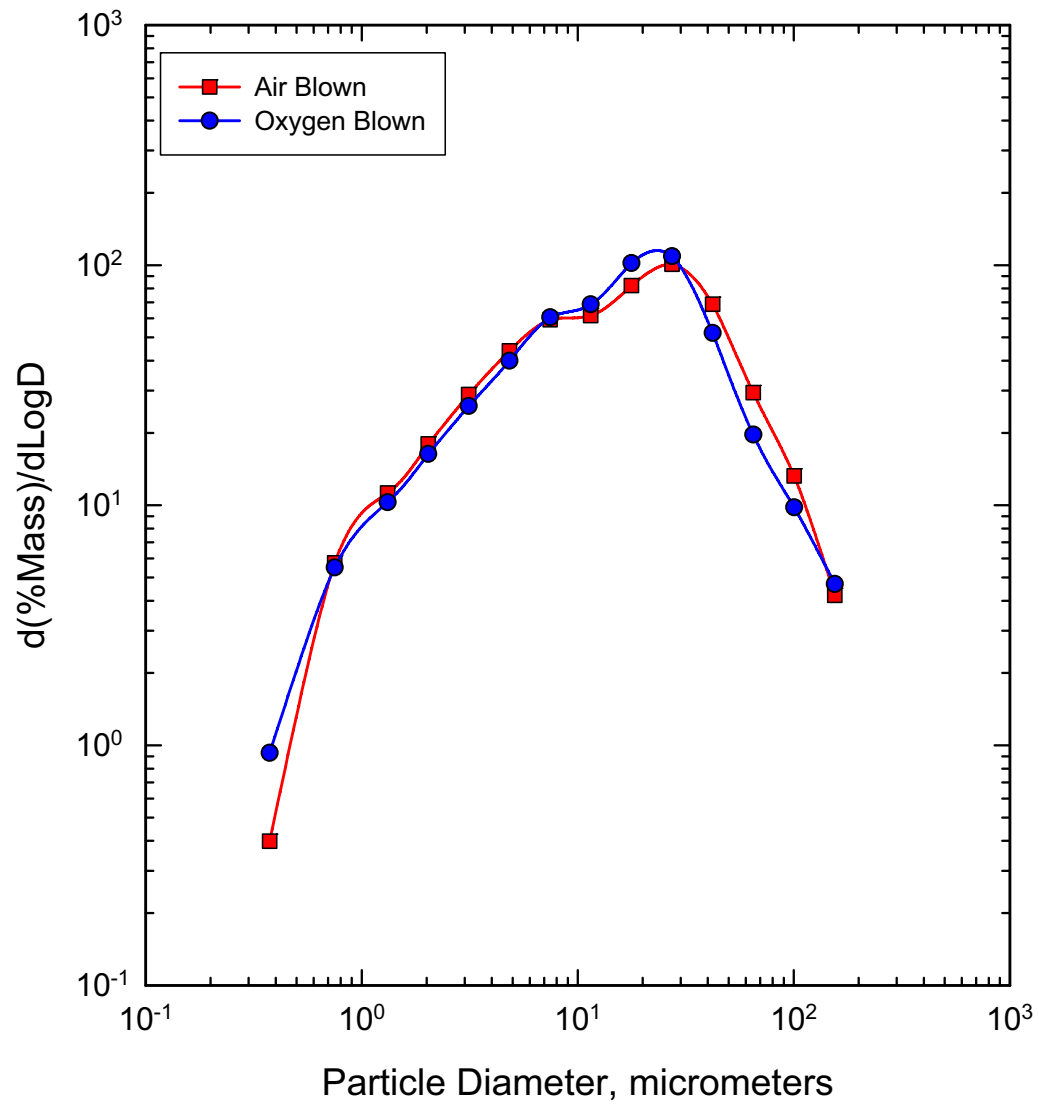


Figure 3.2-6 Comparison of Average PCD Inlet Particle-Size Distributions on Percentage Basis

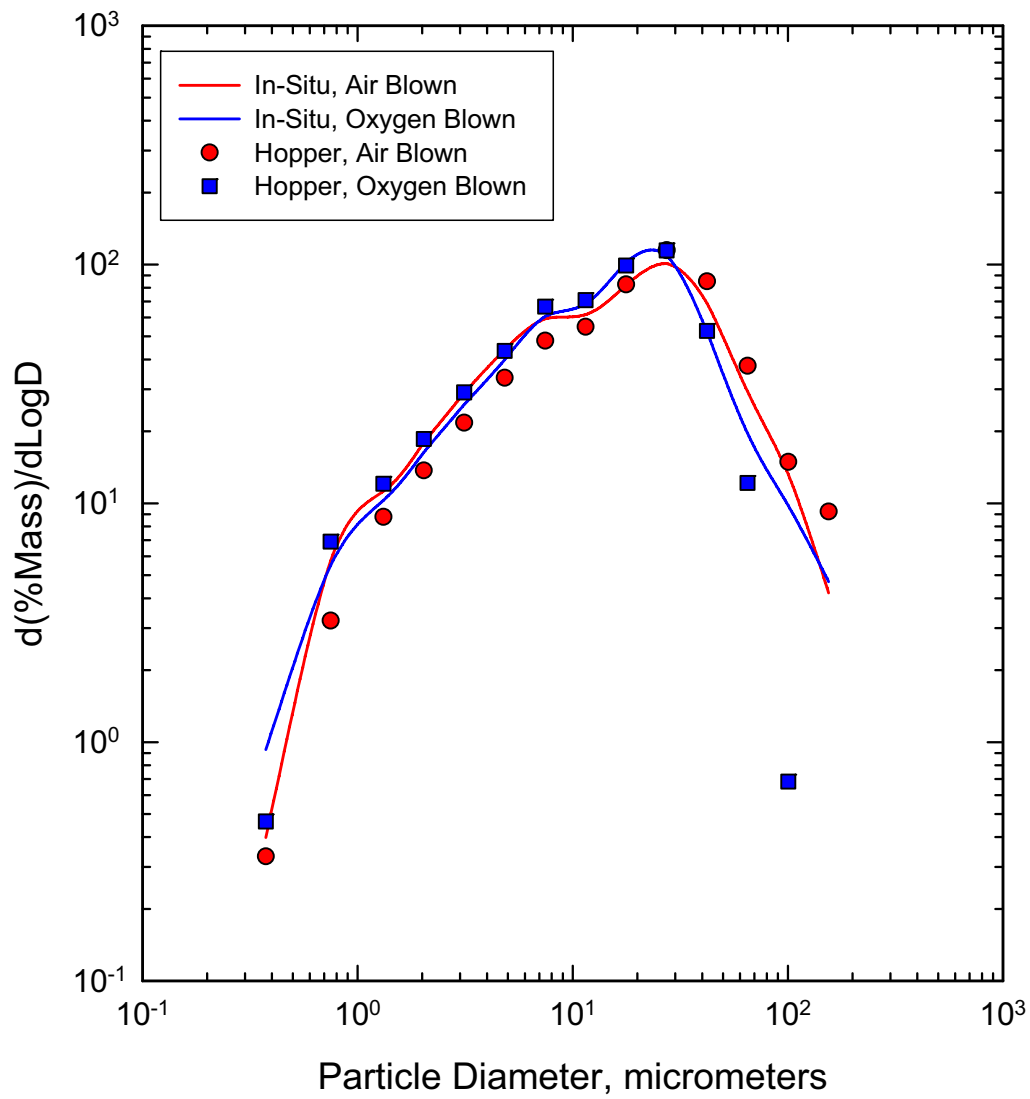


Figure 3.2-7 Comparison of In situ and Hopper Particle-Size Distributions



Figure 3.2-8 Photomicrograph Showing Large Pore Structure of TC15 Residual Cake

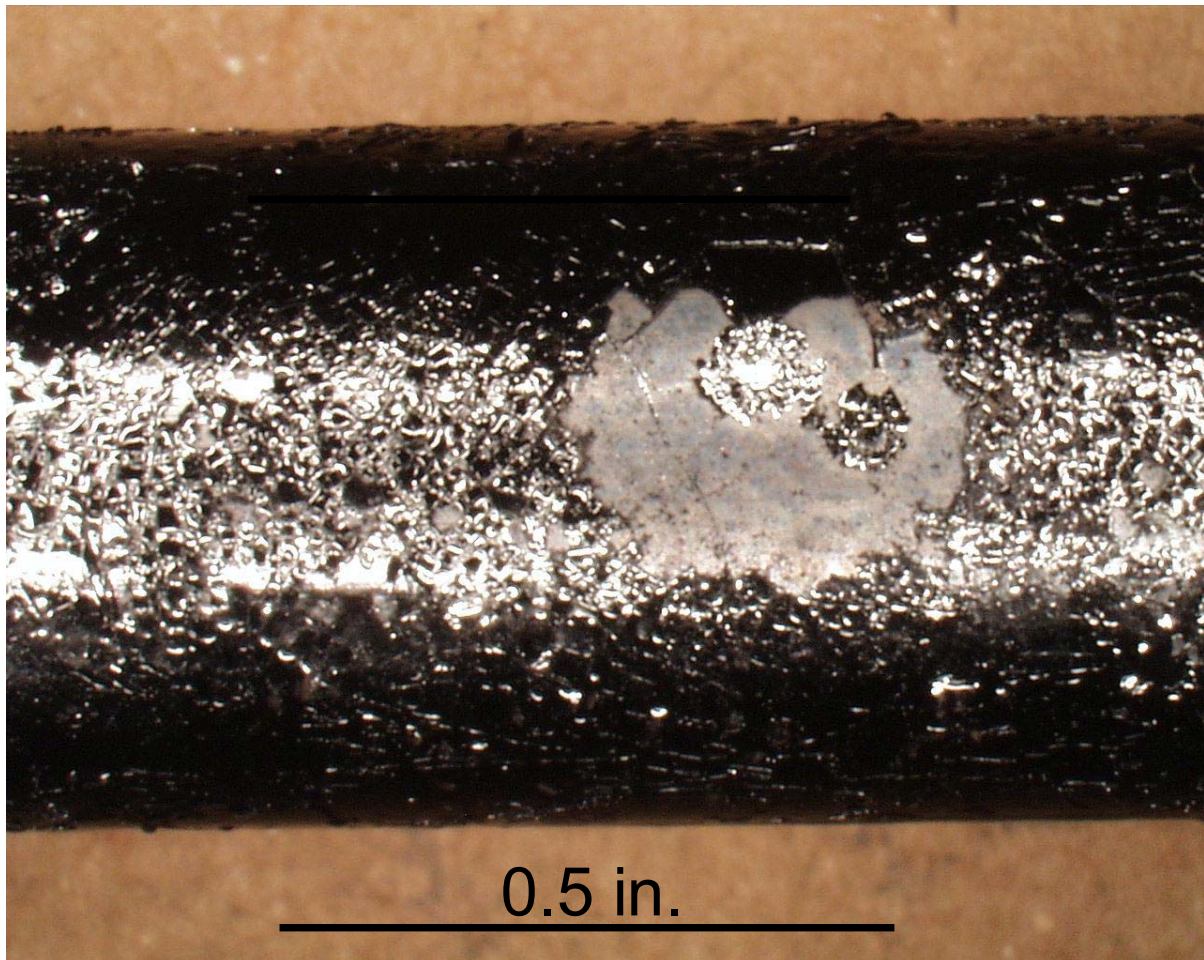


Figure 3.2-9 Photomicrograph Showing Tar Deposition on PCME Probe After TC15

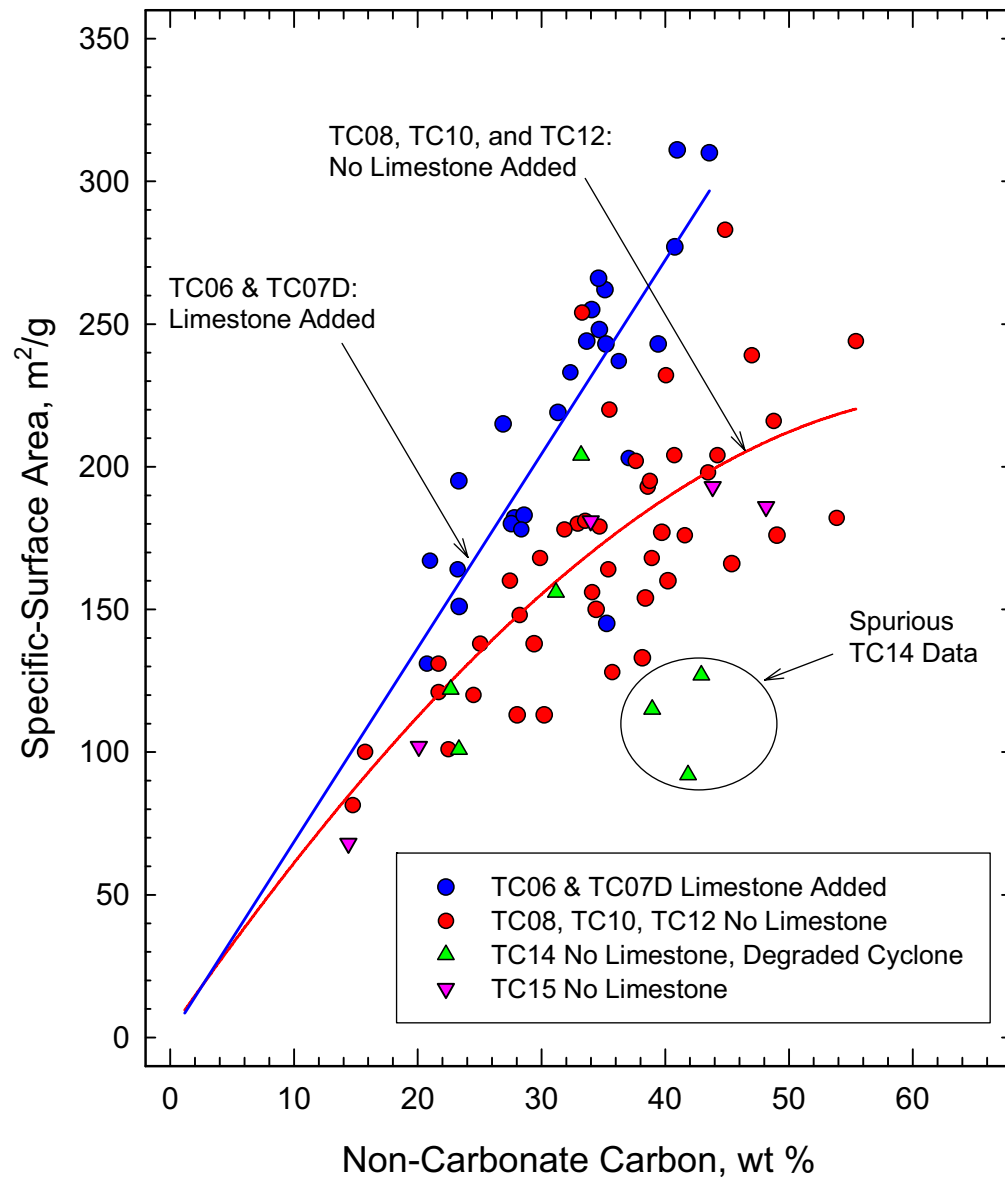


Figure 3.2-10 Specific-Surface Area Versus Carbon Content of In situ Samples

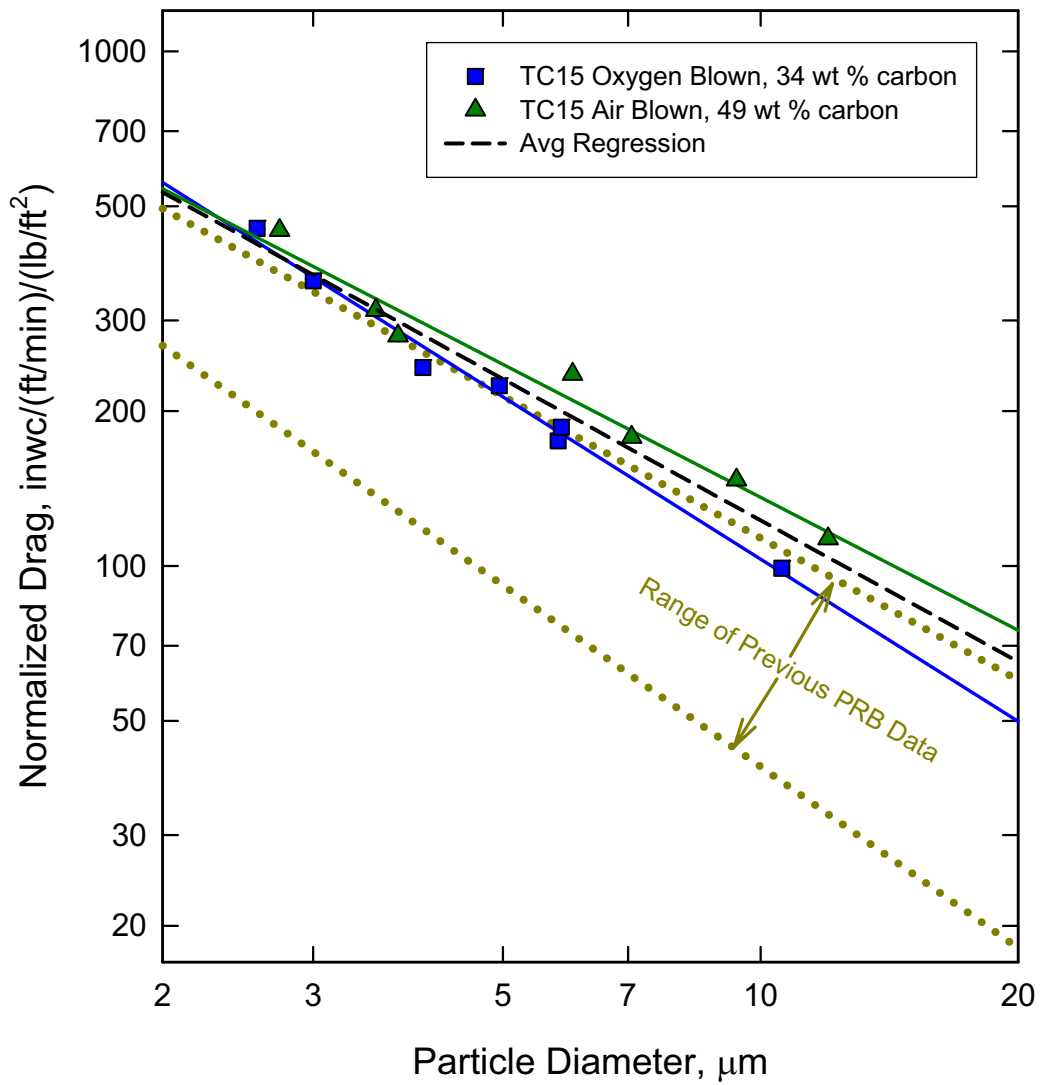


Figure 3.2-11 Laboratory Measurements of TC15 Dustcake Drag Versus Particle Size

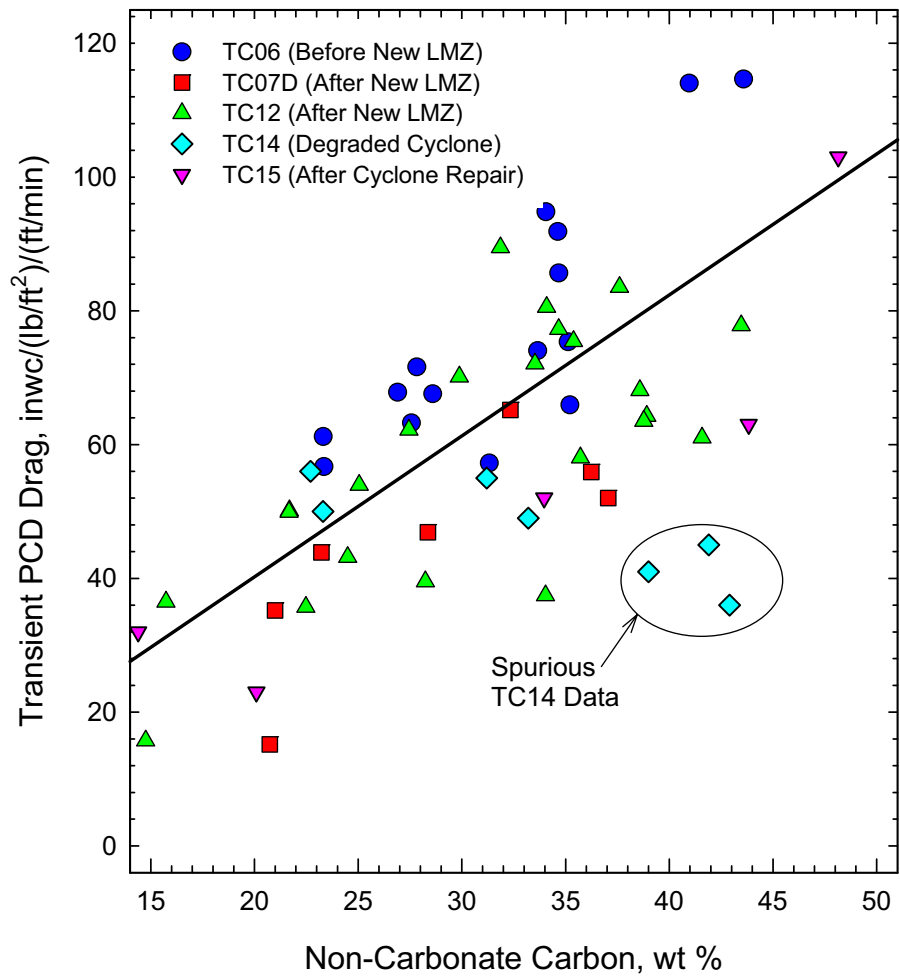


Figure 3.2-12 PCD Transient Drag Versus Carbon Content of In situ Samples

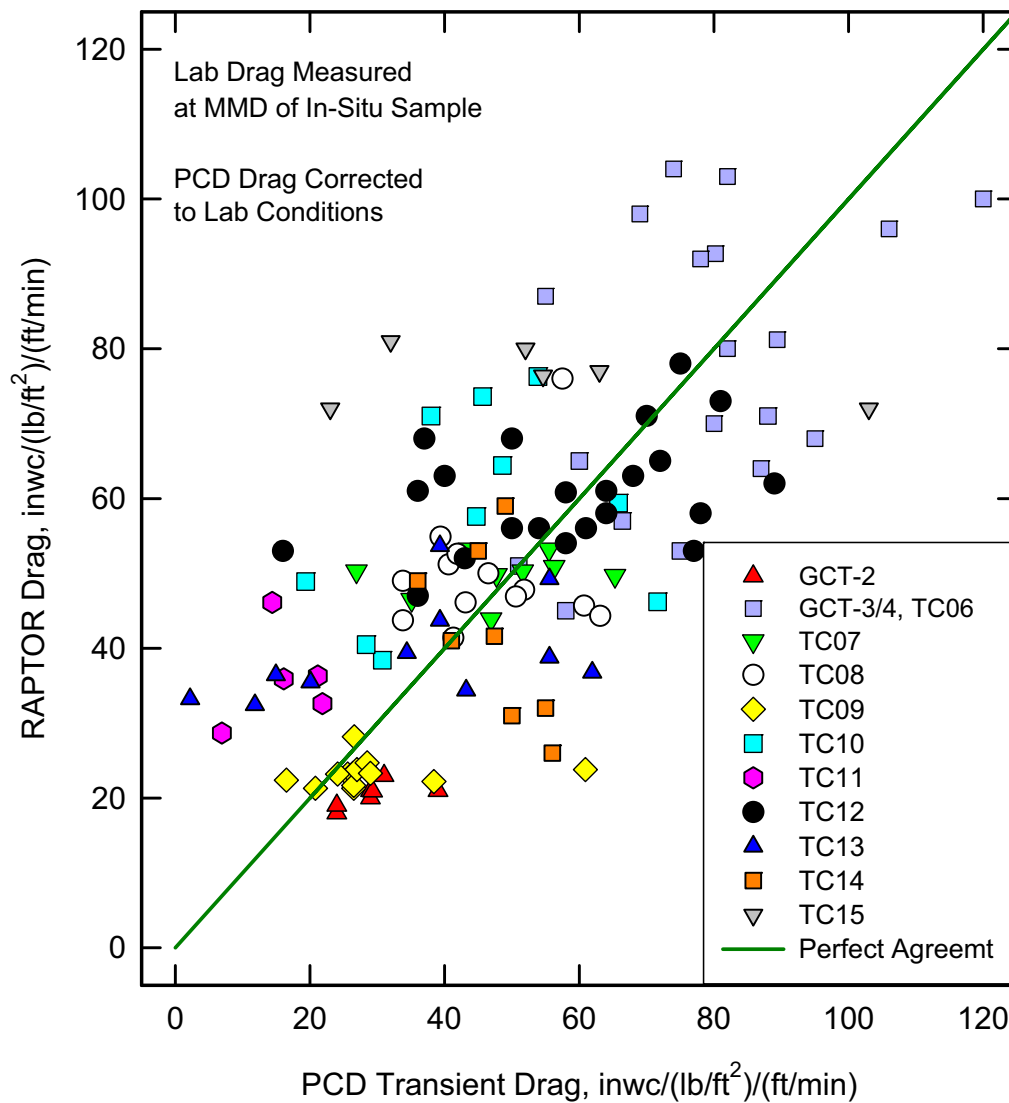


Figure 3.2-13 Comparison of PCD Transient Drag With Laboratory Measurements
(Comparison Based on Individual Runs)

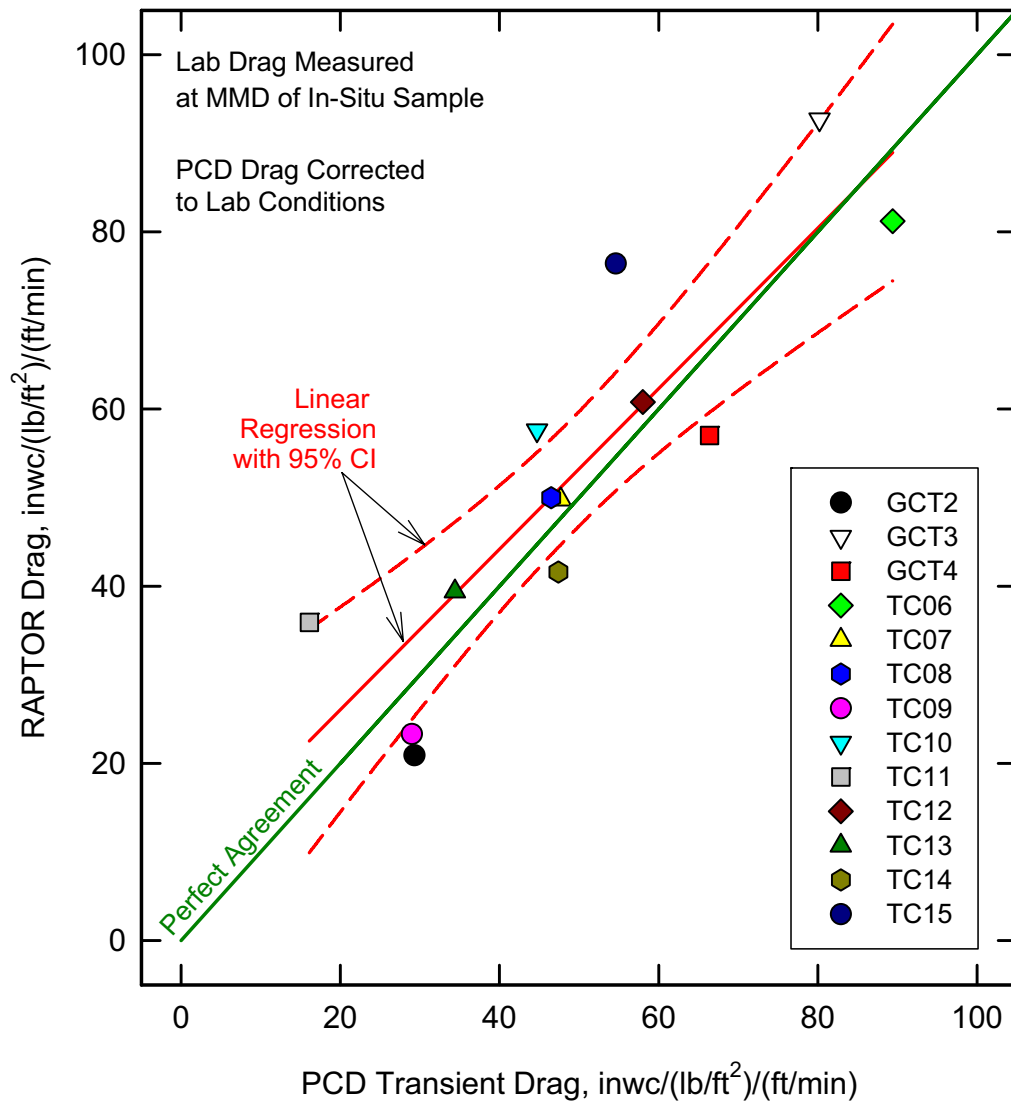


Figure 3.2-14 Comparison of PCD Transient Drag With Laboratory Measurements
(Comparison Based on Average Drag Values)

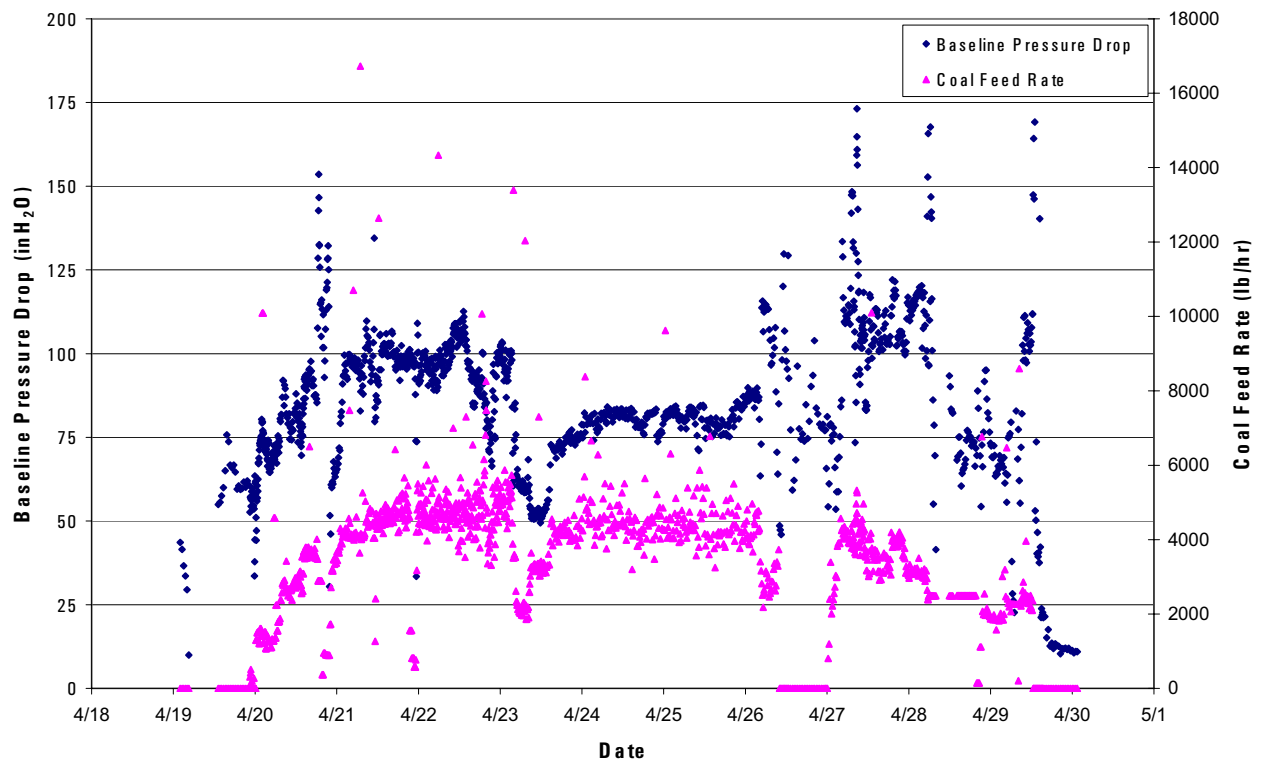


Figure 3.2-15 Baseline Pressure Drop and Coal-Feed Rate During TC15

3.3 OTHER SYSTEMS

3.3.1 Piloted Syngas Burner/Combustion Turbine

The PSB operated on syngas almost 15 hours on April 21 to April 22, 2004, at syngas-flow rates up to 17,000 pph. The unit produced 23.5 MW. The wet raw syngas heating value was between 59 and 64 Btu/scf. As the propane flow rate decreased, the flame remained stable to the point that about 86 percent of the total energy input to the CT came from syngas. A PSB flame temperature greater than 2,800°F gave a reliable indication of approaching flame instability as propane flow was decreased and syngas flow was increased. The flame temperature was maintained above 2,800°F to confirm this finding, and the PSB operated smoothly. When the propane flow decreased below 270 pph, the flame blew out. Before the flame blew out, the stack CO content increased to about 200 to 300 ppm.

The PSB wall temperatures and the combustor noise remained low throughout the run. The improvements to the fuel nozzle thermocouples reduced the frequency and intensity of the temperature spikes that occurred in the previous runs. The size of the thermocouples was larger, and the tips were embedded in the nozzle and covered with a thermal barrier coating. Also contributing to the reduced temperature spikes at the PSB fuel nozzle was an improved control scheme for introducing the main propane, pilot propane, syngas, and purge flows to the PSB. Optimizing these flows appeared to move the PSB flame away from the nozzle.

The PSB was not run on syngas produced during oxygen-blown operation due to circulation issues in the Transport Gasifier.

3.3.2 Primary Cyclone

During the previous test run (TC14), erosion in the primary cyclone caused the gasifier to continuously lose bed material. In TC15, the repaired cyclone had a much higher efficiency, and the solids carryover rate was much lower, despite a high coal-feed rate as shown in [Figure 3.3-1](#). However, the cyclone efficiency was slightly lower than it was originally.

3.3.3 Coal-Feed Systems

The original coal feeder operated well in TC15, supplying about 329 tons of PRB coal to the gasifier over a 9 day period. The developmental coal-feeder fed for a period of about 7 hours. Unfortunately, the deposit that formed during the oxygen nozzle commissioning covered the feed nozzle of the developmental coal feeder and prevented testing during the latter portion of the test run.

3.3.4 Hot Gas Cleanup

The gas cleanup commissioning test run was conducted from April 20 through April 25, 2004. During the testing, the gasifier was operating in air- and oxygen-blown modes using PRB coal. The gas cleanup unit was commissioned for a period of 35 hours of fixed-bed sulfidation

utilizing two Sud-Chemie catalysts (RVS1 and RVSLT1) as the sulfur sorbents. The sorbent bed temperature mostly ranged from 550 to 700°F at pressures 210 psig during air-blown operations and 135 psig in oxygen-blown operations. The minireactor was commissioned for ammonia cracking with simulated bottle gas using Sud-Chemie catalyst (G-117RR). The nominal sorbent properties for fixed-bed desulfurization and ammonia cracking are shown in [Tables 3.3-1](#) and [3.3-2](#). The actual operating parameters for fixed-bed sulfidation in the hot vessels using syngas and ammonia cracking in the minireactor using bottle gas are shown in [Tables 3.3-3](#) and [3.3-4](#).

Test Modifications

The following modifications were completed at the gas cleanup unit during the outage preceding test run TC15.

- The hot vessels (RX700A and RX700B) were rebuilt with stainless steel material and external electrical band heaters were added.
- A new syngas preheater (HX-3) was installed upstream of the hot vessels.
- The new knock-out pot (RX701B) was modified to include a chiller for cooling and steam injection for purging.
- A new stainless steel dirty water tank was added with a steam and nitrogen supply to discharge dirty water to the syngas combustor.
- Two gas analyzer systems, a FTIR and a GC were installed.
- A pressure controller (PIC4944A) was added.
- High temperature electrical heat tracing and insulation was installed to avoid condensation in the process gas lines.
- A new minireactor was installed for the ammonia cracking catalyst test.
- Process lines, instrumentations and controls, and a number of support structures were completely redesigned and installed to achieve reliable operation of hot and cold gas cleanup units.

Test Operation

The major accomplishments and observations in the TC15 gas cleanup test run included the following:

- The gas cleanup unit was operated for 35 hours using two desulfurizer Sud-Chemie sorbents, RVS-1 and RVSLT-1. The gasifier was operating with PRB coal in air- and oxygen-blown modes during the two tests.
- Test 1 was a 19-hour fixed-bed sulfidation test using syngas during air- and oxygen-blown gasifier operation. Sud-Chemie RVS-1 sorbent was installed in the first hot vessel, RX700A. The operating temperature ranged from 550 to 700°F at a pressure of 210 psig during air-blown operations and 650°F at a pressure of 135 psig in oxygen-blown operations. The H₂S concentration at the inlet and outlet of hot vessel, RX700A, varied from 160 to 620 ppm and 10 to 60 ppm. A sulfur loading of 19.2 percent was achieved using RVS-1 sorbent. The temperature and pressure profile and sulfur profile for test 1 are shown in [Figures 3.3-1](#) and [3.3-2](#).
- Test 2 was a 16-hour fixed-bed sulfidation test using syngas during oxygen-blown gasifier operation. Sud-Chemie RVSLT-1 sorbent was installed in the second hot vessel,

RX700B. The operating temperature ranged from 625 to 700°F at pressures 130 to 140 psig during oxygen-blown operations. The H₂S concentration at the inlet and outlet of hot vessel, RX700B varied from 540 to 620 ppm and 10 to 30 ppm. A sulfur loading of 30.4 percent was achieved using RVSLT-1 sorbent. The temperature and pressure profile and sulfur profile for test 2 are shown in [Figures 3.3-3 and 3.3-4](#).

- Intermittent plugging at the restricted orifice in the FTIR gas sample line was observed. The restricted orifice was removed and a needle valve was installed along with a pressure gauge to maintain the sample gas flow.
- Liquid backflow from the cold unit to the gas sample lines was the major problem that affected the FTIR and GC operation.
- Condensation in the process lines and hydrocarbon deposition at the flow control valve downstream of the cold unit caused unreliable gas flow measurement.
- The new minireactor was commissioned while passing a mixture of simulated bottle gas through a nickel-based ammonia cracking catalyst, Sud-Chemie G-117RR. The test unit performed well. The minireactor temperature ranged from 900 to 1,650°F and the pressure ranged from 2 to 10 psig. The mixture of the simulated bottle gas used was NH₃, H₂, and N₂. The sketch of the minireactor unit and the minireactor are shown in [Figures 3.3-5 and 3.3-6](#). The temperature effect on ammonia cracking is shown in [Figure 3.3-7](#).
- Blank tests in the minireactor without catalyst indicated that the nickel-based reactor material and the chrome steel balls employed for preheating the syngas absorbed sulfur and partially decomposed the ammonia. The nickel-based catalyst (G-117RR) was effective in decomposing ammonia when the minireactor temperature exceeded 1,500°F. The catalyst decomposed 98 percent of the ammonia at 1,650°F and 2 to 10 psig. The hydrogen and nitrogen concentrations at the system pressure are the important factors in controlling the exit ammonia concentration. Sulfur compounds, such as H₂S, COS, and CS₂, were injected along with the simulated bottle gas mixture of NH₃, H₂, and N₂ to test the poisoning effect on the ammonia cracking catalyst. No significant degradation of the catalyst was observed as shown in [Figure 3.3-8](#).
- The electrical band heaters around the hot vessels (RX700A/B/C) worked well to control the heating and cooling during testing. A steady vessel temperature up to 1,000°F was maintained during the test run.
- The preheater (HX-3) did not perform as expected. The syngas temperature inlet to the hot vessels was limited to 600°F primarily due to high heat loss in the outlet piping at the lower syngas flow.
- The knock-out pot (RX701B) was commissioned along with water chiller (HX701). Hydrocarbon crystals were formed in the knock-out pot while circulating chiller water at 40°F but were removed using 50 psig steam purging.
- The dirty water tank (TK700) effectively collected process condensate from the water scrubber (RX701) and discharged to the syngas combustor using 50 psig steam and/or nitrogen.
- Gas sampling systems (FTIR and GC with FPD) were commissioned using syngas. The GC primarily measured H₂S, COS, and CS₂ while the FTIR analyzed for water vapor, ammonia, and various hydrocarbons. Gas sampling was intermittent and limited due to a condensation problem in the sample line during oxygen-blown operation.

- The pressure controller (PIC4944A) performed well to maintain 15 psig back pressure. However, condensation and hydrogen crystal deposition interrupted smooth operation of the pressure controller.
- The heat tracing in the gas cleanup unit was well maintained at around 600°F to eliminate condensation in the process lines.

Table 3.3-1

Fixed-Bed Desulfurization Sorbents in Hot Vessels

Sud-Chemie Sorbent	RVS-1	RVSLT-1
Chemical Composition	wt%	wt%
Zinc Oxide	40 - 60	50 - 70
Calcium Sulfate	15 - 25	15 - 30
Calcium Oxide	5 - 10	5 - 15
Nickel Oxide	5 - 15	-
Bentonite	5 -15	5 - 15
Silica, Quartz	<5	<5
Physical Properties		
Shape	Spheres	Spheres
Size	3 - 4 mm	3 - 4 mm
Density	60 - 85 lb/ft ³	60 - 85 lb/ft ³

Table 3.3-2

Fixed-Bed Ammonia Cracking Catalyst in Minireactor

Sud-Chemie Sorbent	G-117RR
Chemical Composition	wt%
Magnesium Oxide	75 - 90
Nickel Oxide	5 - 15
Calcium Oxide	1 - 5
Aluminum Oxide	1 - 5
Physical Properties	
Shape	Rings
Size	3 - 4 mm
Density	55 - 75 lb/ft ³

Table 3.3-3

Actual Operating Parameters for Fixed-Bed Sulfidation in Hot Vessels Using Syngas

Set Point	SP-1	SP-2
Gasifier Operation	Air /O ₂ Blown	O ₂ Blown
Coal Type	PRB	PRB
Reactor	RX700A	RX700B
Reactor Size	5.187" ID x 5' Ht	5.187" ID x 5' Ht
Reactor Material	310 SS	310SS
Sorbent	Sud-Chemie RVS-1	Sud-Chemie
RVSLT-1		
Sorbent bed mass, lb	2	2
Sorbent bed height, in	2.3	2.3
Syngas flow rate, lb/hr	45 - 3	12
Pressure, psig	210 - 130	135
Temperature, °F	550 - 700	650
Space Velocity, hr ⁻¹	24,000 - 1,700	6,700
Inlet H ₂ S, ppm	160 – 620	580

Table 3.3-4

Actual Operating Parameters for Fixed-Bed Ammonia Cracking in Minireactor Using Bottle Gas

Minireactor	RX-301
Minireactor Size	1.5" ID x 4' Ht
Sorbent	Sud-Chemie G-117RR
Sorbent bed mass, lb	0.25
Sorbent bed height, in	5
Syngas flow rate, lb/hr	1 - 1.5
Pressure, psig	2 - 13
Temperature, °F	900 - 1,650
Space Velocity, hr ⁻¹	2,000 - 3,000
NH ₃ inlet, ppm	2,500 - 3,000
NH ₃ conversion, %	80 - 98 (from 1100 to 1650°F)

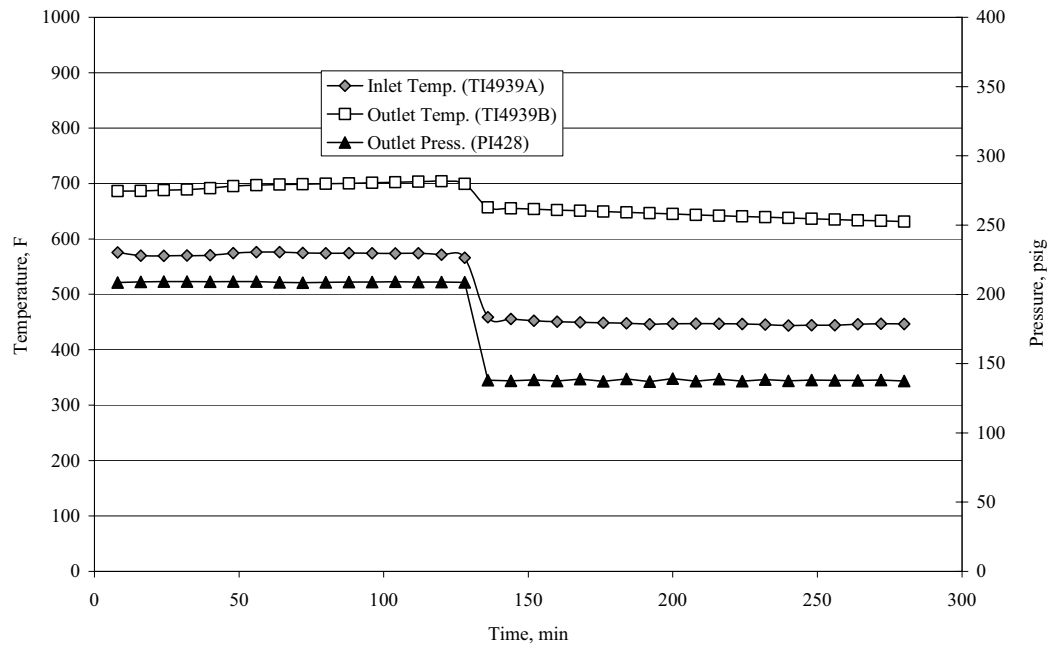


Figure 3.3-1 RX700A Temperature and Pressure Profile: RVS-1 Sorbent

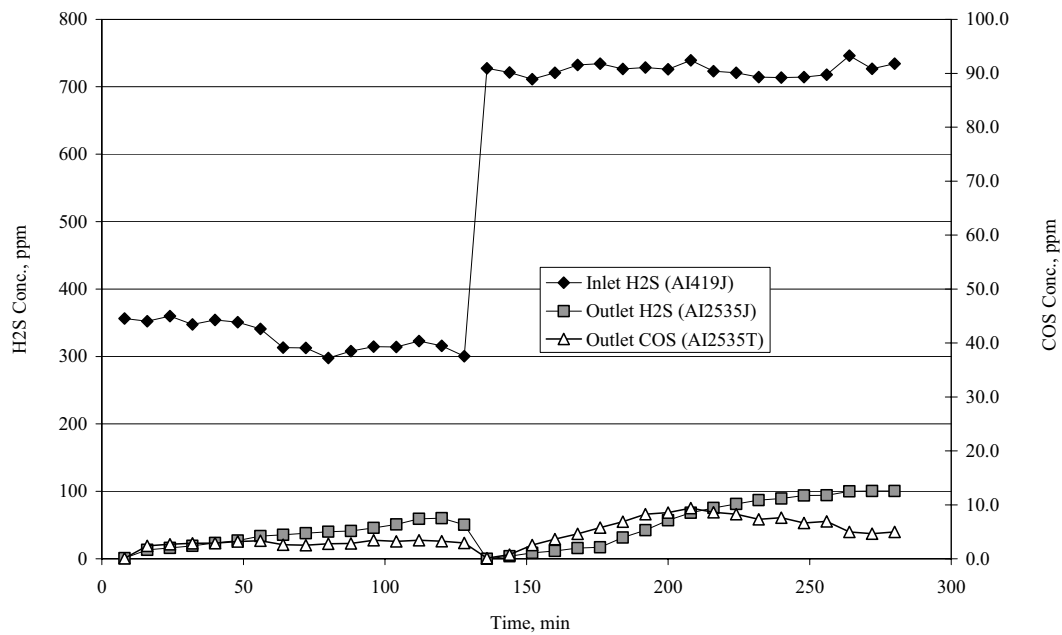


Figure 3.3-2 RX700A Sulfur Profile: RVS-1 Sorbent

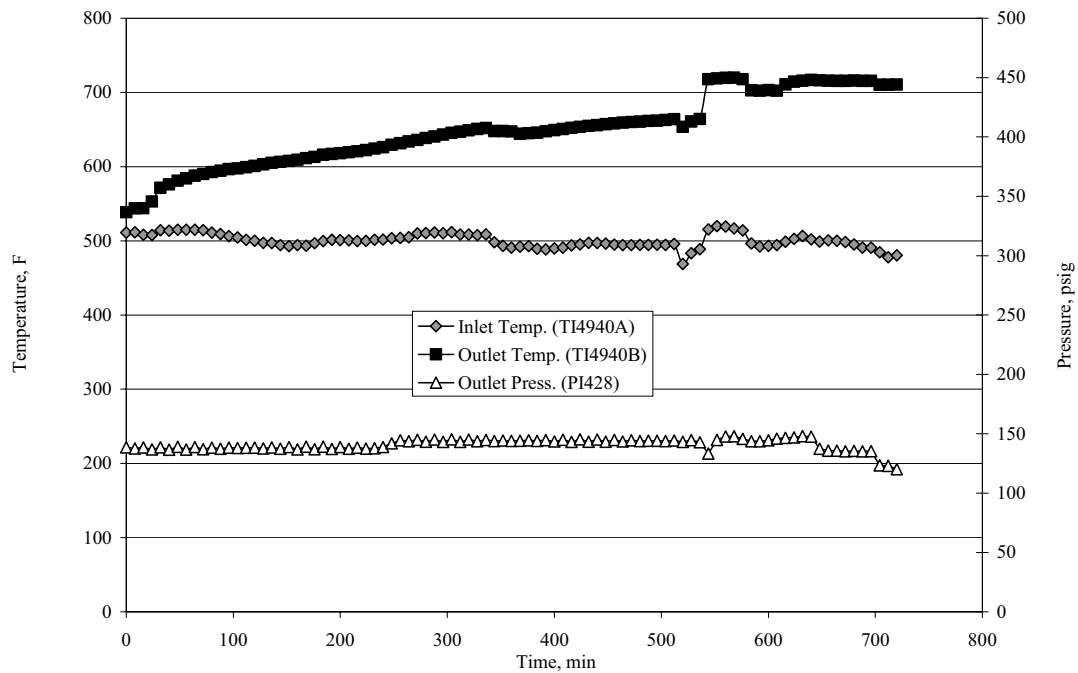


Figure 3.3-3 RX700B Temperature and Pressure Profile: RVSLT-1 Sorbent

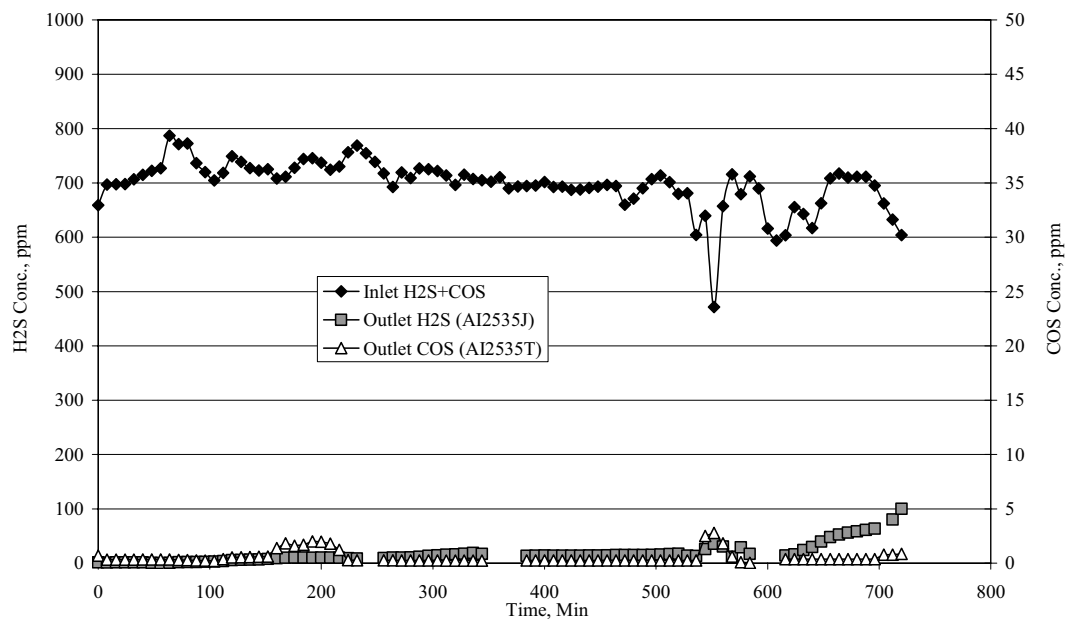


Figure 3.3-4 RX700B Sulfur Profile: RVSLT-1 Sorbent

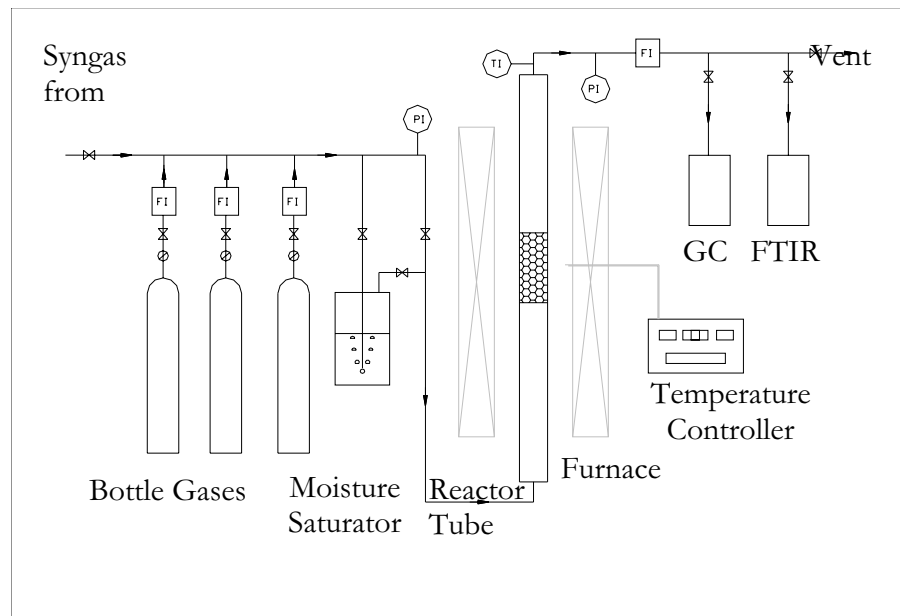


Figure 3.3-5 Sketch of Minireactor Unit

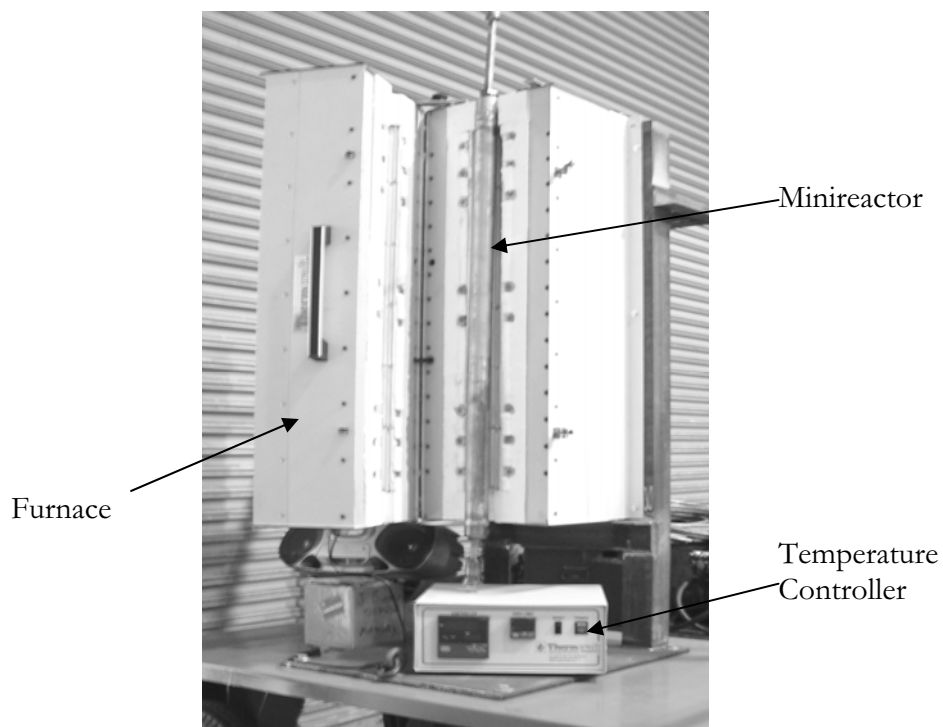


Figure 3.3-6 Minireactor Unit

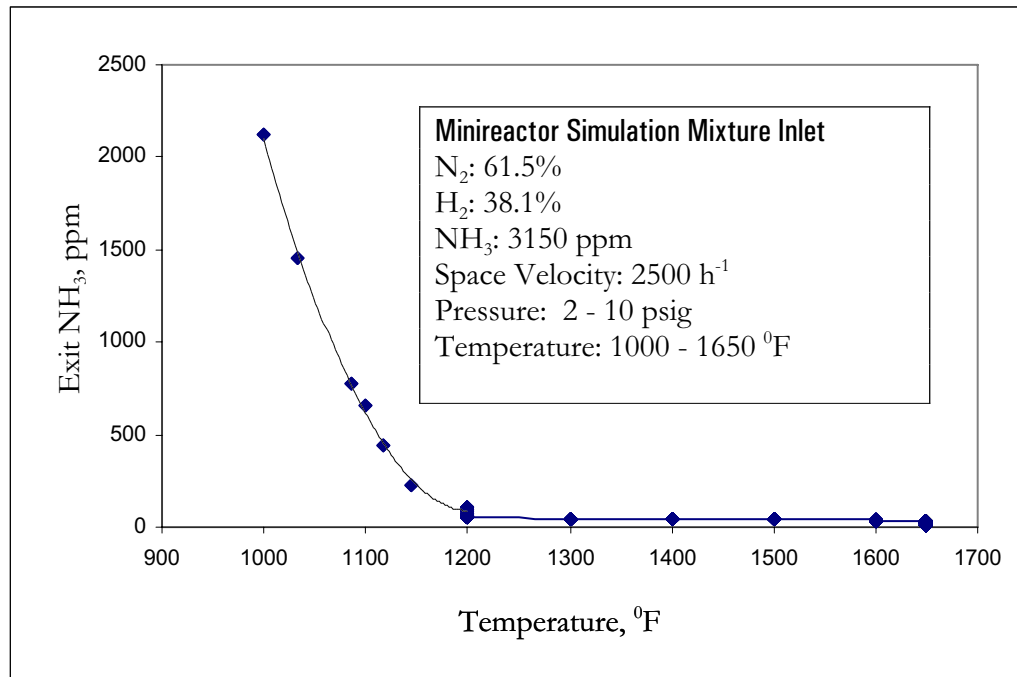


Figure 3.3-7 Exit Ammonia Versus Temperature Profile: G-117RR Ni-Catalyst

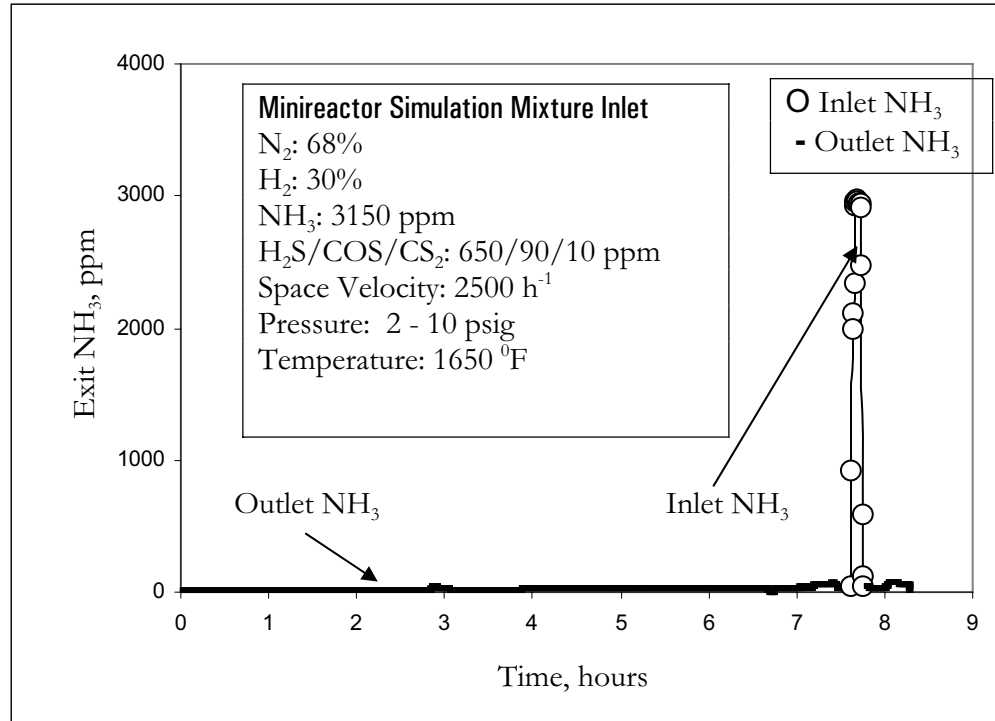


Figure 3.3-8 Simulation Test of Ammonia Cracking: G-117RR Ni-Catalyst With Sulfur Injection

4.0 CONCLUSIONS

In conclusion, TC15 was a moderately successful test run accumulating 200 hours of on-coal operation using PRB coal. Approximately 91 hours of operation were in oxygen-blown gasification and the balance in air-blown gasification and air-to-oxygen transition. The agglomeration that formed in the mixing zone due to a temperature excursion prevented oxygen distribution testing from occurring and shortened the duration of the test run, however, the gas cleanup tests, PSB operation, and PCD failsafe and face velocity tests were successfully conducted as scheduled. The major highlights included:

- The repaired and slightly modified cyclone inlet had a higher efficiency resulting in a lower solids carryover rate at a higher coal-feed rate.
- The CFAD system operated reliably and appears to be a viable alternative to using lock hoppers for pressurized solids removal.
- The PCD operated stably throughout TC15 with no filter element failures or gasification ash bridging. The reduction in filter surface area resulted in a slightly higher baseline pressure drop. The outlet particle loading was initially elevated, but dropped below 0.1 ppmw, the lower limit of detection, for the remainder of the test run. Failsafe testing with hot gasification ash injection was successfully completed on the CeraMem ceramic failsafe in which the failsafe plugged quickly.
- The hot gas cleanup system was sufficient for cleaning the syngas to levels required for operating a solid oxide fuel cell.
- The PSB operated on syngas for around 15 hours, at syngas flow rates up to 17,000 pph showing that the turbine can operate reliably on syngas with a low BTU content (85 to 150 Btu/scf).

APPENDIX A1 OPERATION HISTORY

Conversion of the Transport Reactor train to gasification mode of operation was performed from May to September 1999. The first gasification test run, GCT1, was a 233-hour test run to commission the Transport Gasifier and to characterize the limits of operational parameter variations. GCT1 was started on September 9, 1999, with the first part completed on September 15, 1999. The second part of GCT1 was started on December 7, 1999, and was completed on December 15, 1999. This test run provided the data necessary for preliminary analysis of gasifier operations and for identification of necessary modifications to improve equipment and process performance. Five different feed combinations of coal and sorbent were tested to gain a better understanding of the gasifier solids collection system efficiency.

GCT2, a 218-hour characterization test run, was started on April 10, 2000, and was completed on April 27, 2000. Additional data was taken to analyze the effect of different operating conditions on gasifier performance and operability. A blend of several PRB coals was used with Longview limestone from Alabama. In the outage following GCT2, the Transport Gasifier was modified to improve the operation and performance of the gasifier solids collection system. The most fundamental change was the addition of the loop seal underneath the primary cyclone.

GCT3 was a 184-hour characterization with the primary objective to commission the loop seal. A hot solids circulation test (GCT3A) was started on December 1, 2000, and was completed on December 15, 2000. After a 1-month outage to address maintenance issues with the main air compressor, GCT3 was continued. The second part of GCT3 (GCT3B) was started on January 20, 2001, and was completed on February 1, 2001. During GCT3B, a blend of several PRB coals was used with Bucyrus limestone from Ohio. The loop seal performed well, allowing much higher solids circulation rates and higher syngas heating values. Also, the improved collection efficiency of the cyclone resulted in lower relative solids loading to the PCD and higher carbon conversion.

GCT4, a 242-hour characterization test run, was started on March 7, 2001, and was completed on March 30, 2001. A blend of several PRB coals with Bucyrus limestone from Ohio was used. More experience was gained with the loop seal operations and additional data was collected to better understand gasifier performance.

TC06, a 1,025-hour test campaign, was started on July 4, 2001, and was completed on September 24, 2001. A blend of several PRB coals with Bucyrus limestone from Ohio was used. Due to its length and stability of operation, the TC06 test run provided valuable data necessary to analyze long-term gasifier operations and to identify necessary modifications to improve equipment and process performance, as well as, satisfying the goal of many thousands of hours of filter element exposure.

TC07, a 442-hour test campaign, was started on December 11, 2001, and was completed on April 5, 2002. A blend of several PRB coals and a bituminous coal from the Calumet mine in Alabama were tested with Bucyrus limestone from Ohio. Due to operational difficulties with the gasifier (stemming from instrumentation problems), the unit was taken offline several times. PCD operations were relatively stable considering the numerous gasifier upsets.

TC08 was a 365-hour test campaign to commission the gasifier in the oxygen-blown mode of operation. TC08 started on June 9, 2002, and was completed on June 29, 2002. A blend of several PRB coals were tested in air-blown, enriched-air, and oxygen-blown modes of operation. The transition from different modes of operation was smooth and it was demonstrated that the transition from air to oxygen could be made within 15 minutes. Both gasifier and PCD operations were stable during the test run.

TC09 was a 309-hour test campaign to characterize the gasifier and PCD operations in air- and oxygen-blown modes of operations using a bituminous coal from the Sufco mine in Utah. TC09 was started on September 3, 2002, and was completed on September 26, 2002. Both gasifier and PCD operations were stable during the test run.

TC10 was a 416-hour test campaign to conduct long-term tests to evaluate the gasifier and PCD operations in oxygen-blown mode of operations using a blend of several PRB coals. TC10 was started on November 16, 2002, and was completed on December 18, 2002. Despite problems with the coal mills, coal feeder, pressure tap nozzles, and the standpipe, the gasifier did experience short periods of stability during oxygen-blown operations. During these periods, the syngas quality was high. During TC10, over 609 tons of PRB subbituminous coals were gasified.

TC11 was a 192-hour test campaign to conduct short-term tests to evaluate the gasifier and PCD operations in air- and oxygen-blown modes of operations using Falkirk lignite from North Dakota. TC11 was started on April 7, 2003, and was completed on April 18, 2003. During TC11, the lignite proved difficult to feed due to difficulties in the mill operation as a result of the high moisture content in the fuel. However, the gasifier operated well using lignite, with high circulation rates, riser densities, and steady temperature profiles. Consequently, the temperature distribution in both the mixing zone and the riser was more uniform than in any previous test run, varying less than 10°F throughout the gasifier.

TC12 was a 733-hour test campaign to conduct short-term tests to evaluate the gasifier and PCD operations in air- and oxygen-blown modes of operations using a blend of several PRB coals. TC12 was started on May 16, 2003, and was completed on July 14, 2003. A primary focus for TC12 was the commissioning of a new gas clean-up system and operating a fuel cell on syngas derived from the Transport Gasifier. The fuel cell system and gas clean-up system performed well during the testing.

TC13 was a 501-hour test campaign to conduct short-term tests to evaluate gasifier, PSB, and PCD operations in air-blown mode of operations using a blend of several PRB coals, as well as, to conduct short-term tests to evaluate gasifier and PCD operations using two different types of lignite from the Freedom mine in North Dakota. One type of lignite had high ash sodium content, while the other types had low ash sodium content. TC13 was started on September 30, 2003, and was completed on November 2, 2003. The syngas-to-PSB testing lasted for a total of about 6 hours. While successful, the hydraulic system on the turbine cranking motor failed and prevented further PSB testing. The low sodium lignite testing went well, but lowering the gasifier temperature to below 1,500°F was necessary to prevent ash agglomeration with the high sodium lignite.

TC14 was a 214-hour test campaign to conduct short-term tests to evaluate the gasifier, PSB, and PCD operations in air- and oxygen-blown modes of operations using a blend of several PRB coals. TC14 began on February 16, 2004, and ended on February 28, 2004. The syngas-to-PSB testing lasted for a total of about 17 hours at syngas flow rates up to 17,000 pph, contributing about 82 percent of the total energy to the PSB. The CFAD unit was commissioned during TC14. The new system worked well and operated for 190 hours. The gasifier operation was smooth, with the exception of the decrease in the primary cyclone efficiency which caused the gasifier to continuously lose bed material.

TC15, the subject of this report, was a 200-hour test campaign to conduct short-term tests to evaluate the gasifier, PSB, and PCD operations in air- and oxygen-blown modes of operation using a blend of several PRB coals. TC15 began on April 19, 2004, and ended on April 29, 2004. The syngas-to-PSB testing lasted for approximately 15 hours at syngas-flow rates up to 17,000 pph, contributing about 86 percent of the total energy to the PSB. The gasifier experienced stable operations in air-blown mode and less stable operations in oxygen-blown mode due to poor solids circulation. A primary focus of TC15 was to commission and test modifications made to the syngas cleanup system. The system was effective in reducing the sulfur content and achieving an absorption capacity between 19 and 30 percent.

Figure A1-1 gives a summary of operating test hours achieved with the Transport Reactor at the PSDF.

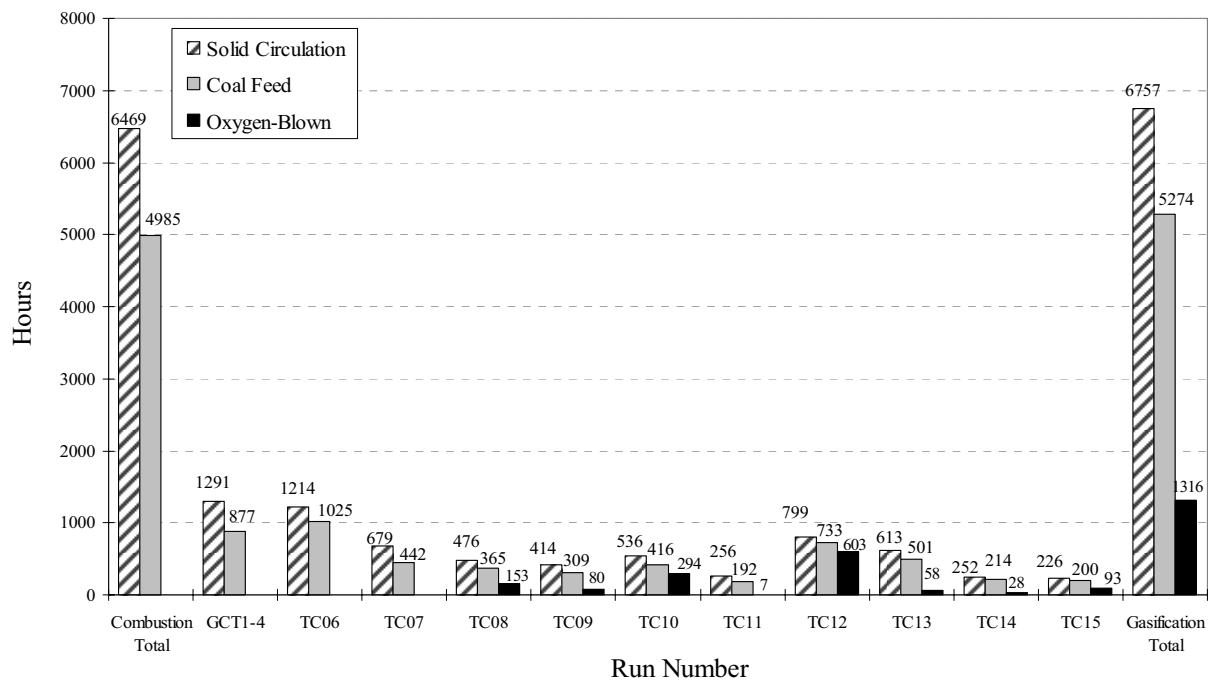


Figure A1-1 Operating Hours Summary for the Transport Reactor Train

APPENDIX A2 EQUIPMENT LIST

Major Equipment in the Transport Gasifier Train

TAG NAME	DESCRIPTION
BR0201	Reactor Start-Up Burner
BR0401	Atmospheric Syngas Combustor Oxidizer
BR0452	Piloted Syngas Burner
BR0602	AFBC Start-Up Burner
CO0201	Main Air Compressor
CO0451	Turbine Air Compressor
CO0601	AFBC Air Compressor
CY0201	Primary Cyclone in the Reactor Loop
CY0207	Disengager in the Reactor Loop
CY0601	AFBC Cyclone
DR0402	Steam Drum
DY0201	Feeder System Air Dryer
FD0206	Spent Solids Screw Cooler
FD0210	Coal-Feeder System
FD0220	Sorbent Feeder System
FD0502	Fines Screw Cooler
FD0510	Spent Solids Transporter System
FD0520	Fines Transporter System
FD0530	Spent Solids Feeder System
FD0602	AFBC Solids Screw Cooler
FD0610	AFBC Sorbent Feeder System
FL0301	PCD — Siemens Westinghouse
FL0401	Compressor Intake Filter
GN0451	Turbine Generator
GT0451	Gas Turbine
HX0202	Primary Gas Cooler
HX0204	Transport Air Cooler
HX0402	Secondary Gas Cooler
HX0405	Compressor Feed Cooler
HX0540	CFAD Collection Drum/Heat Exchanger
HX0601	AFBC Heat Recovery Exchanger
ME0540	Heat Transfer Fluid System
RX0201	Transport Reactor
SI0602	Spent Solids Silo
SU0601	Atmospheric Fluidized-Bed Combustor (AFBC)

Major Equipment in the Balance of Plant (Page 1 of 3)

TAG NAME	DESCRIPTION
B02920	Auxiliary Boiler
B02921	Auxiliary Boiler – Superheater
CL2100	Cooling Tower
C02201A-D	Service Air Compressor A-D
C02202	Air-Cooled Service Air Compressor
C02203	High-Pressure Air Compressor
C02601A-C	Reciprocating N ₂ Compressor A-C
CR0104	Coal and Sorbent Crusher
CV0100	Crushed Feed Conveyor
CV0101	Crushed Material Conveyor
DP2301	Baghouse Bypass Damper
DP2303	Inlet Damper on Dilution Air Blower
DP2304	Outlet Damper on Dilution Air Blower
DY2201A-D	Service Air Dryer A-D
DY2202	Air-Cooled Service Air Compressor Air Dryer
DY2203	High-Pressure Air Compressor Air Dryer
FD0104	MWK Coal Transport System
FD0105	FW Coal Transport System
FD0111	MWK Coal Mill Feeder
FD0112	FW Coal Mill Feeder
FD0113	Sorbent Mill Feeder
FD0140	Coke Breeze and Bed Material Transport System
FD0154	MWK Limestone Transport System
FD0810	Ash Unloading System
FD0820	Baghouse Ash Transport System
FL0700	Baghouse
FN0700	Dilution Air Blower
H00100	Reclaim Hopper
H00105	Crushed Material Surge Hopper
H00252	Coal Surge Hopper
H00253	Sorbent Surge Hopper
HT2101	MWK Equipment Cooling Water Head Tank
HT2103	SCS Equipment Cooling Water Head Tank
HT0399	60-Ton Bridge Crane
HX2002	MWK Steam Condenser
HX2003	MWK Feed Water Heater

Major Equipment in the Balance of Plant (Page 2 of 3)

TAG NAME	DESCRIPTION
HX2004	MWK Subcooler
HX2103A	SCS Cooling Water Heat Exchanger
HX2103B	FW Cooling Water Heat Exchanger
HX2103C	MWK Cooling Water Heat Exchanger
LF0300	Propane Vaporizer
MC3001-3017	MCCs for Various Equipment
ME0700	MWK Stack
ME0814	Dry Ash Unloader for MWK Train
ML0111	MWK Coal Mill
ML0112	FW Coal Mill
ML0113	Sorbent Mill for Both Trains
PG0011	Oxygen Plant
PG2600	Nitrogen Plant
PU2000A-B	MWK Feed Water Pump A-B
PU2100A-B	Raw Water Pump A-B
PU2101A-B	Service Water Pump A-B
PU2102A-B	Cooling Tower Make-Up Pump A-B
PU2103A-D	Circulating Water Pump A-D
PU2107	SCS Cooling Water Make-Up Pump
PU2109A-B	SCS Cooling Water Pump A-B
PU2110A-B	FW Cooling Water Pump A-B
PU2111A-B	MWK Cooling Water Pump A-B
PU2300	Propane Pump
PU2301	Diesel Rolling Stock Pump
PU2302	Diesel Generator Transfer Pump
PU2303	Diesel Tank Sump Pump
PU2400	Fire Protection Jockey Pump
PU2401	Diesel Fire Water Pump #1
PU2402	Diesel Fire Water Pump #2
PU2504A-B	Waste Water Sump Pump A-B
PU2507	Coal and Limestone Storage Sump Pump
PU2700A-B	Demineralizer Forwarding Pump A-B

Major Equipment in the Balance of Plant (Page 3 of 3)

TAG NAME	DESCRIPTION
PU2920A-B	Auxiliary Boiler Feed Water Pump A-B
SB3001	125-V DC Station Battery
SB3002	UPS
SC0700	Baghouse Screw Conveyor
SG3000-3005	4160-V, 480-V Switchgear Buses
SI0101	MWK Crushed Coal Storage Silo
SI0102	FW Crushed Coal Storage Silo
SI0103	Crushed Sorbent Storage Silo
SI0111	MWK Pulverized Coal Storage Silo
SI0112	FW Pulverized Coal Storage Silo
SI0113	MWK Limestone Silo
SI0114	FW Limestone Silo
SI0810	Ash Silo
ST2601	N ₂ Storage Tube Bank
TK2000	MWK Condensate Storage Tank
TK2001	FW Condensate Tank
TK2100	Raw Water Storage Tank
TK2300A-D	Propane Storage Tank A-D
TK2301	Diesel Storage Tank
TK2401	Fire Water Tank
XF3000A	230/4.16-kV Main Power Transformer
XF3001B-5B	4160/480-V Station Service Transformer No. 1-5
XF3001G	480/120-V Miscellaneous Transformer
XF3010G	120/208 Distribution Transformer
XF3012G	UPS Isolation Transformer
VS2203	High-Pressure Air Receiver

APPENDIX A3 MASS AND ENERGY BALANCES

Material balances are useful in checking the accuracy and consistency of data, as well as, determining periods of operation where the data is suitable for model development and commercial plant design. Total material balances for each operating period are given in [Figure A3-1](#) which compare the total mass in and the total mass out. The overall material balance was good, with all of the relative differences at ± 10 percent. The relative difference (relative error) is defined as the Transport Gasifier feeds minus the products divided by the feeds ($\{\text{In-Out}\}/\text{In}$).

The main contributors to the material balance are the syngas-flow rate (12,400 to 22,600 pph), the air-flow rate (0 to 13,000 pph), the oxygen-flow rate (0 to 2,600 pph), the steam-flow rate (300 to 4,000 pph), the nitrogen-flow rate (5,500 to 6,700 pph), and the coal-feed rate (1,700 to 4,400 pph). Although higher than in previous test runs, the sand-feed rate (0 to 110 pph) contributed only a small amount to the overall mass balance. All but two of the air-blown operating periods had higher overall mass flow rates than the oxygen-blown operating periods due to additional nitrogen from air fed to the Transport Gasifier.

The TC15 Transport Gasifier energy balance is shown in [Figure A3-2](#) with standard conditions chosen to be a pressure of 1.0 atmosphere and a temperature of 80°F. As shown in the figure, the TC15 energy balances were mostly within ± 10 percent error with the exception of a few that fell within ± 11 percent error. The energy entering the gasifier consisted of the coal, air, and steam fed to the Transport Gasifier. The nitrogen, oxygen, and sand fed to the gasifier were considered to be at standard conditions (80°F) and, hence, had zero enthalpy. The nitrogen and oxygen feeds actually entered the gasifier at a higher temperature than standard conditions, but compared to the other feed enthalpies, this neglected input energy is insignificant. Since the amount of solids removed from the standpipe was negligible, the energy exiting the gasifier consisted of only the syngas and PCD solids. The analysis used the lower heating value of the coal, the PCD solids, and the syngas.

The energy of the syngas was determined at the Transport Gasifier primary cyclone exit. Since the total syngas flow measurement is located downstream of the PCD, 320 pph of nitrogen that flowed to the PCD inlet and outlet particulate sampling trains was subtracted from the exit flow rate to determine the actual syngas rate from the cyclone. The sensible enthalpy of the syngas was determined by the overall gas heat capacity from the syngas compositions and using gas heat capacities information. The syngas and PCD solids energy consists of latent and sensible heat. The heat loss from the Transport Gasifier was estimated to be 3.5 MBtu/hr.

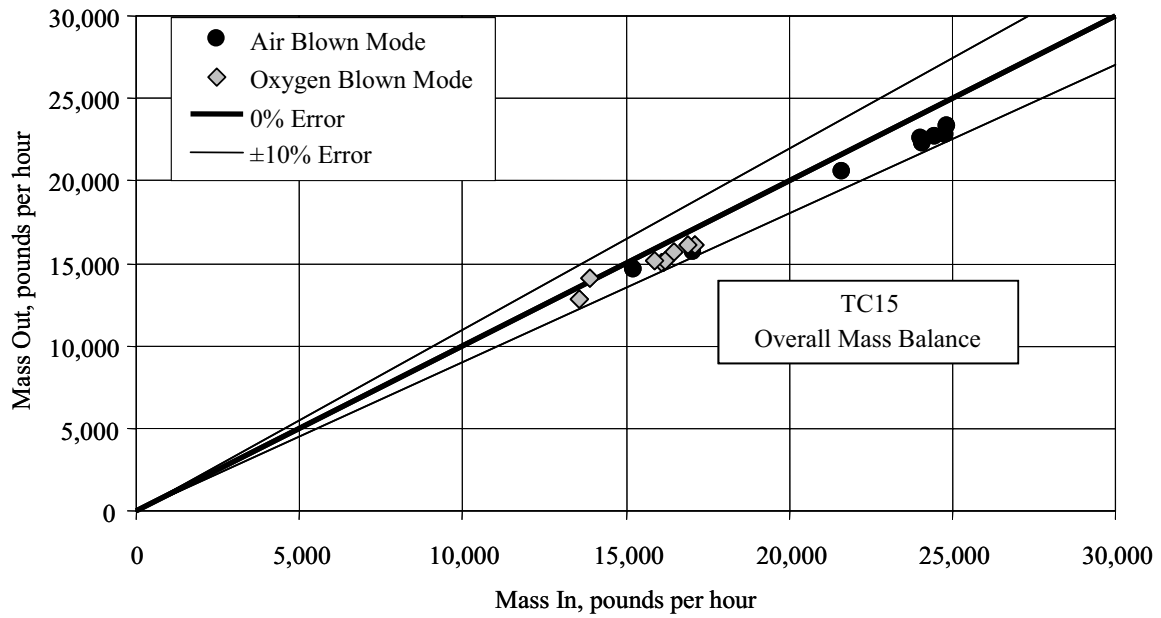


Figure A3-1 Mass Balance

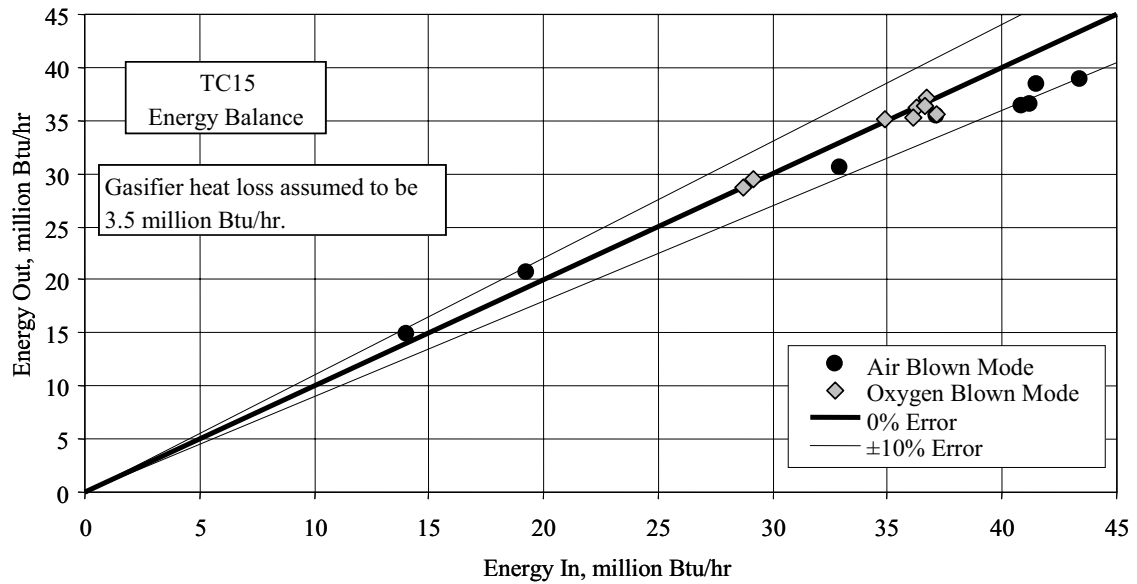


Figure A3-2 Energy Balance

APPENDIX A4 OPERATING GRAPHS

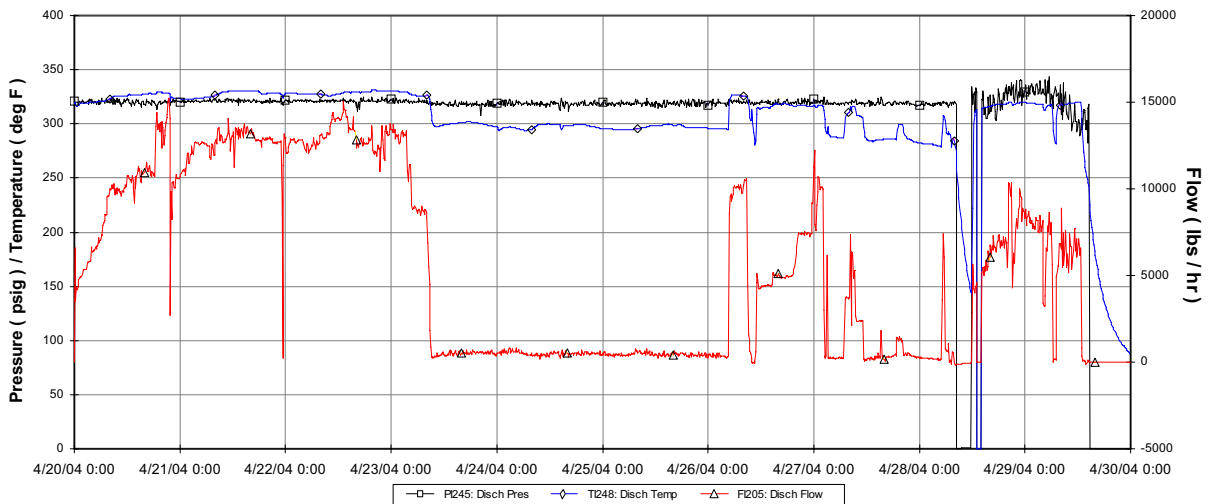


Figure A4-1 C00201 System Profile

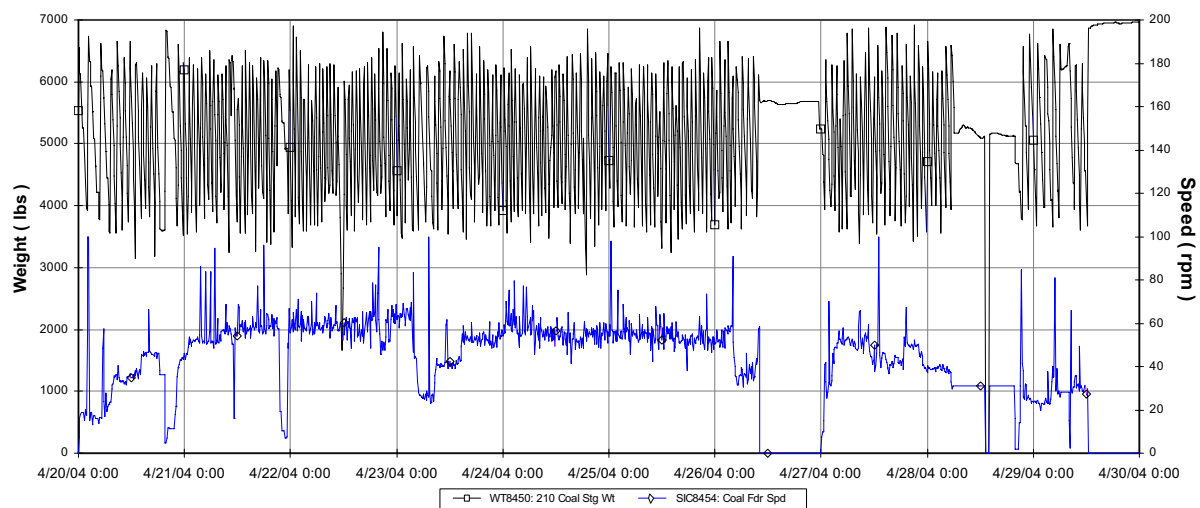


Figure A4-2 Coal-Feed System Profile

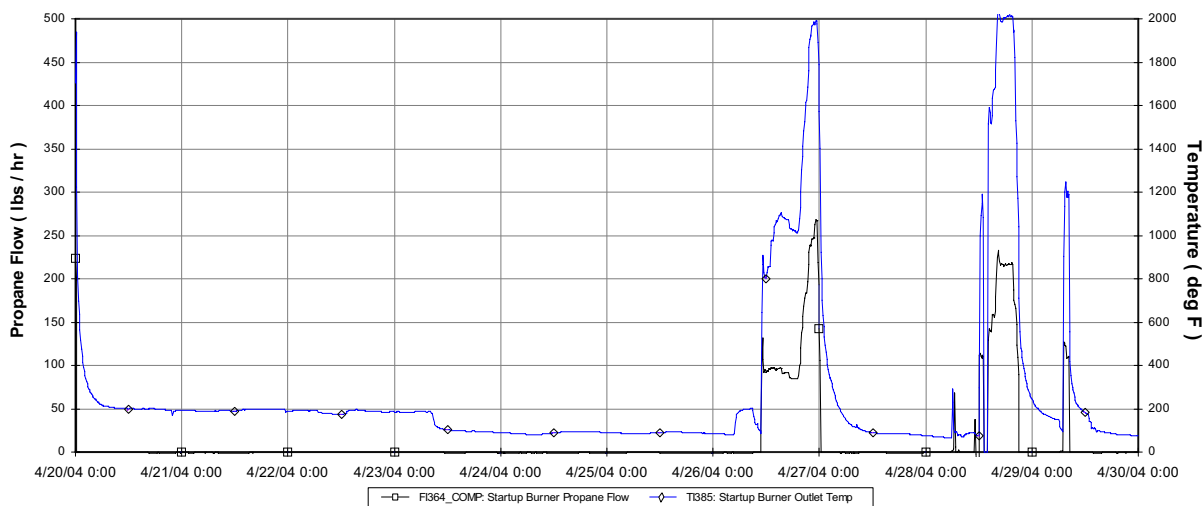


Figure A4-3 Start-Up Burner Flow/Temperature

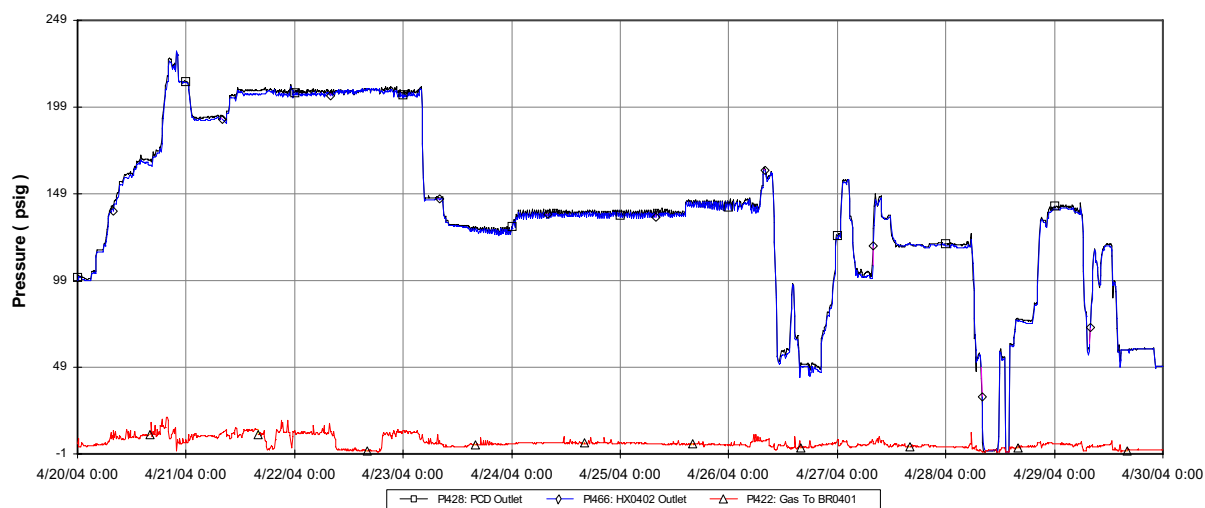


Figure A4-4 System Pressures Downstream of PCD

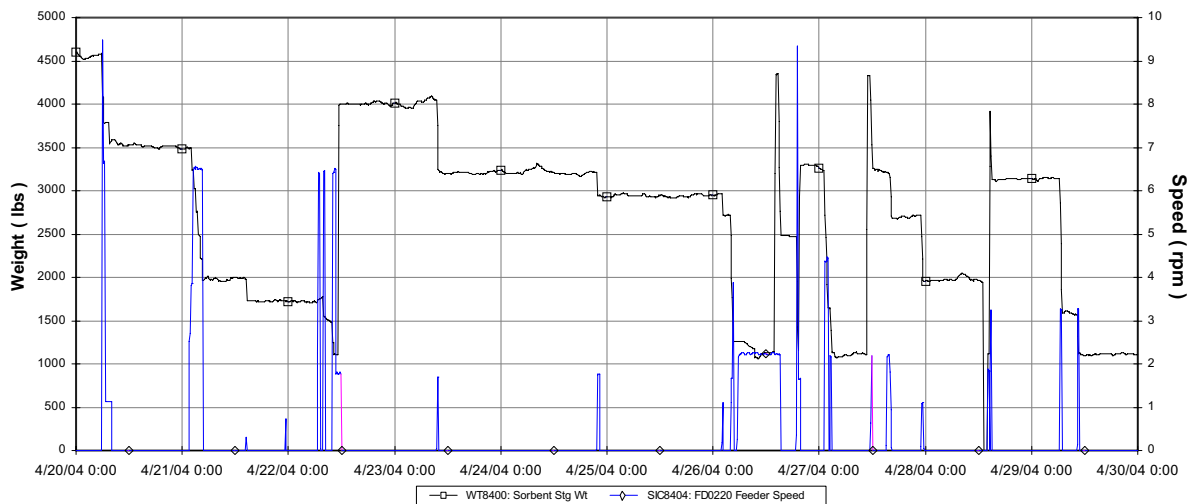


Figure A4-5 Sorbent Feed System Profile

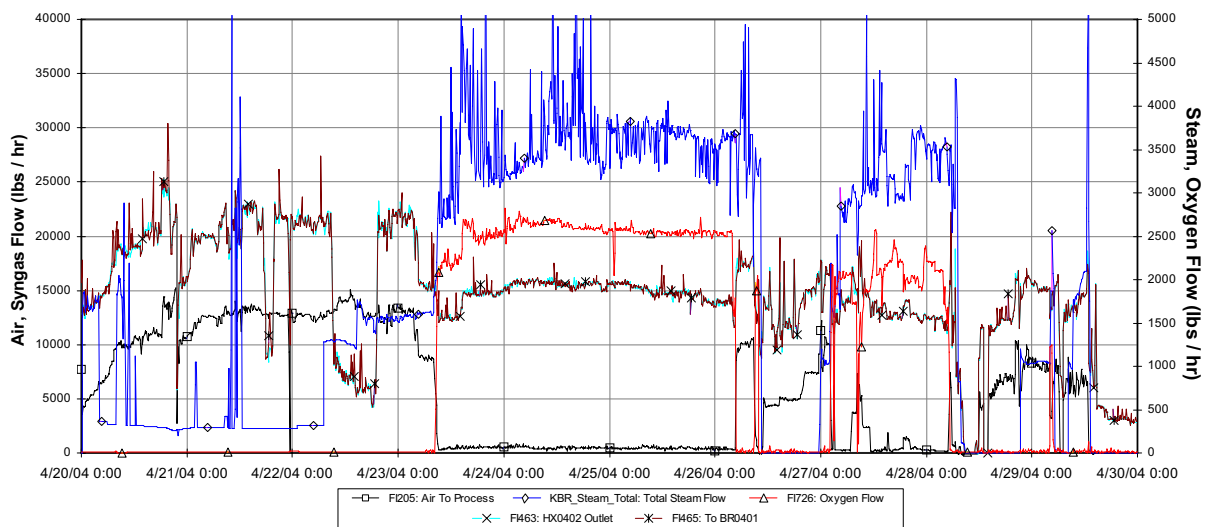


Figure A4-6 Total Gas In/Out Flow Rates

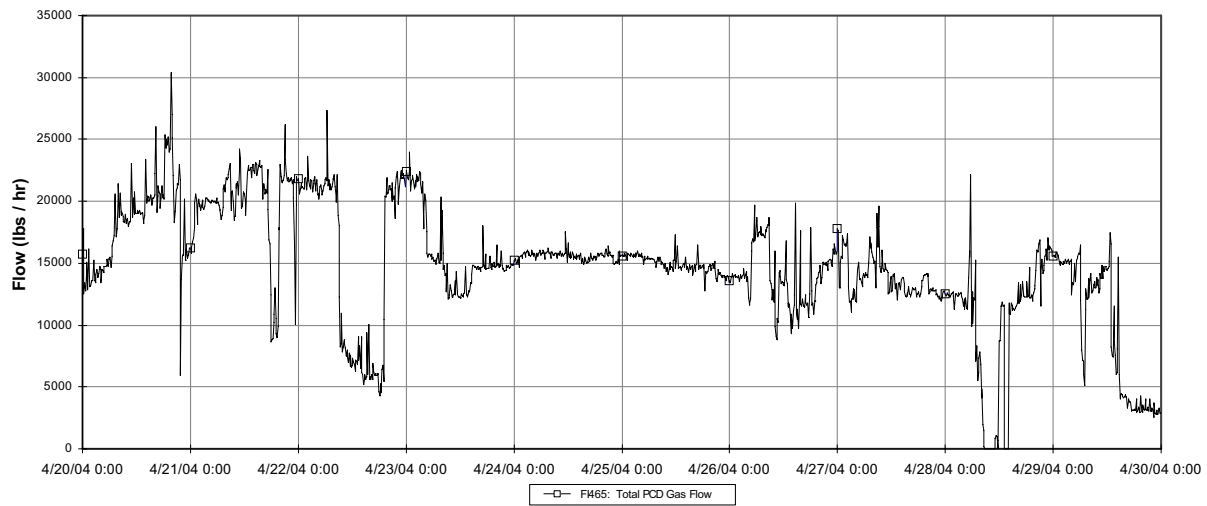


Figure A4-7 Gas Flow Through PCD

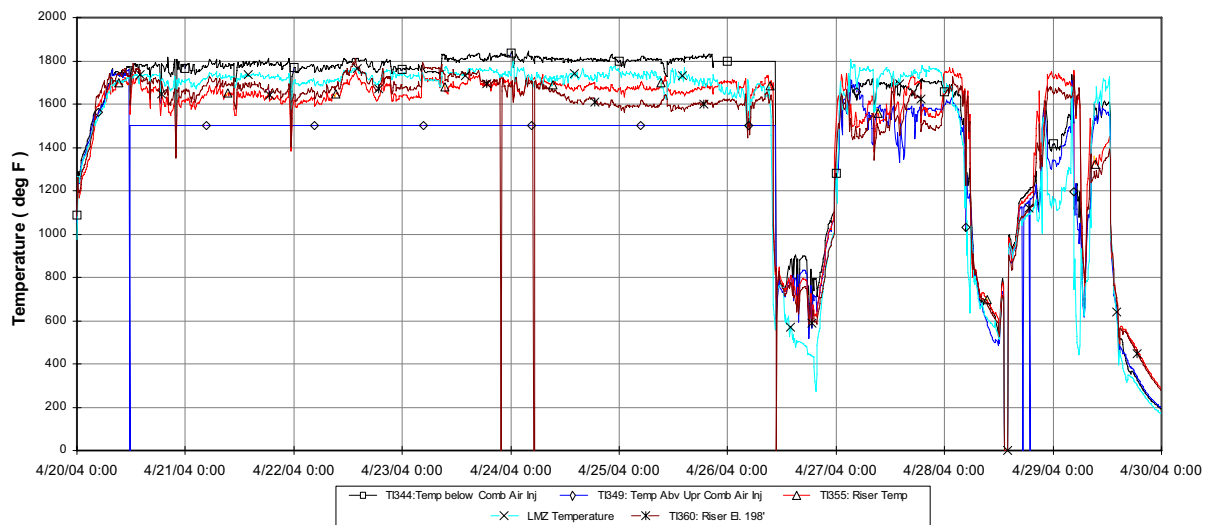


Figure A4-8 Gasifier Mixing Zone and Riser Temperatures

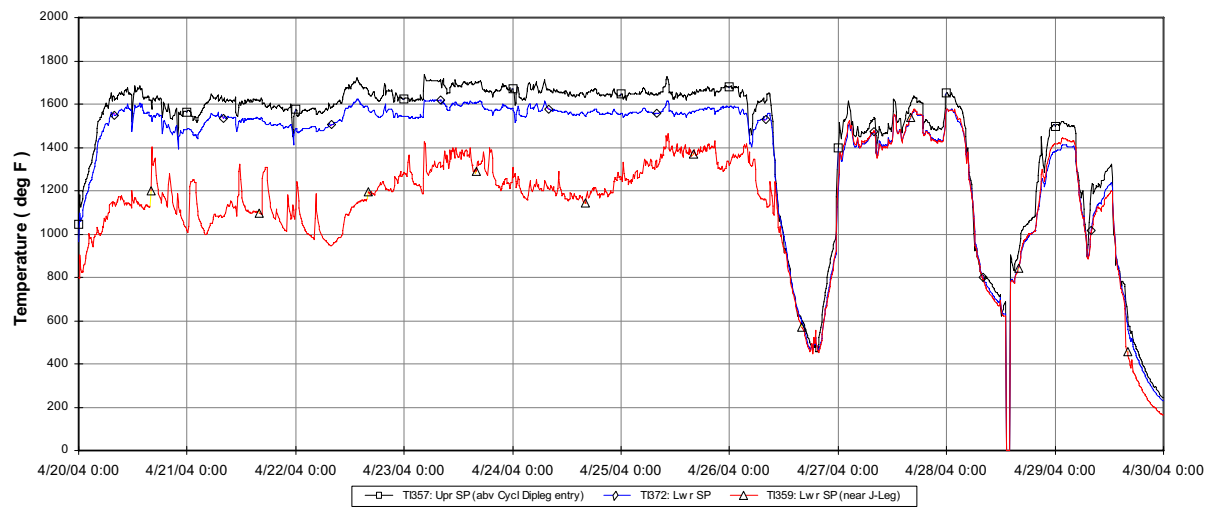


Figure A4-9 Standpipe Temperatures

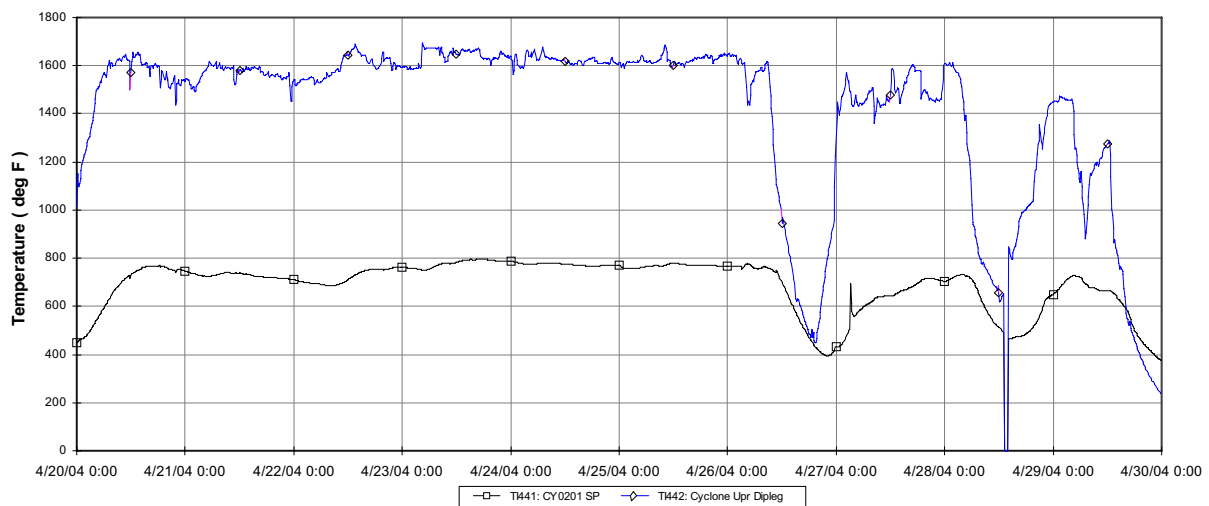


Figure A4-10 Loop Seal Temperatures

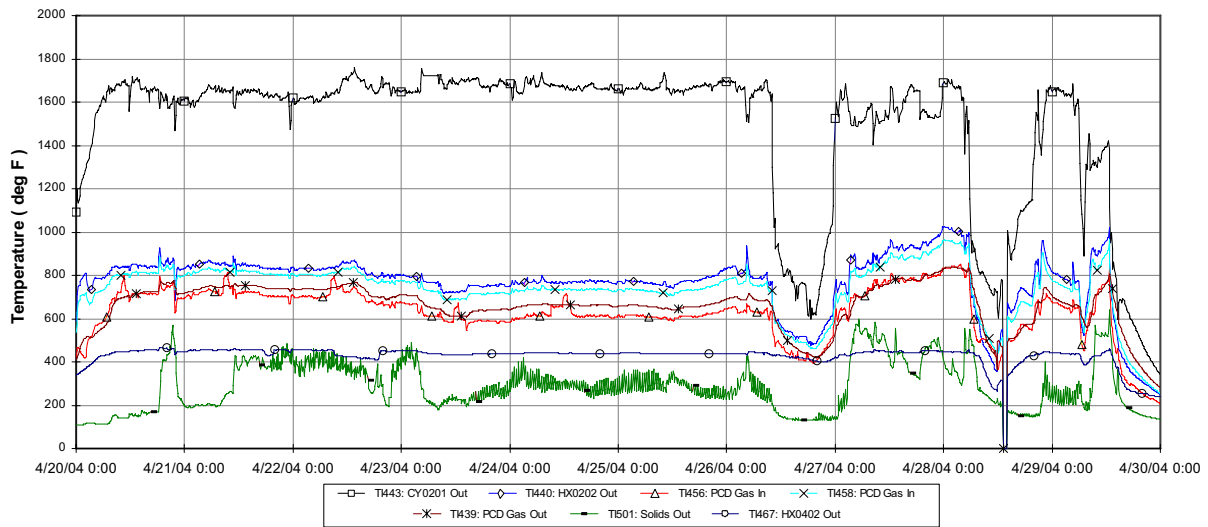


Figure A4-11 Temperature Profile Downstream of Gasifier

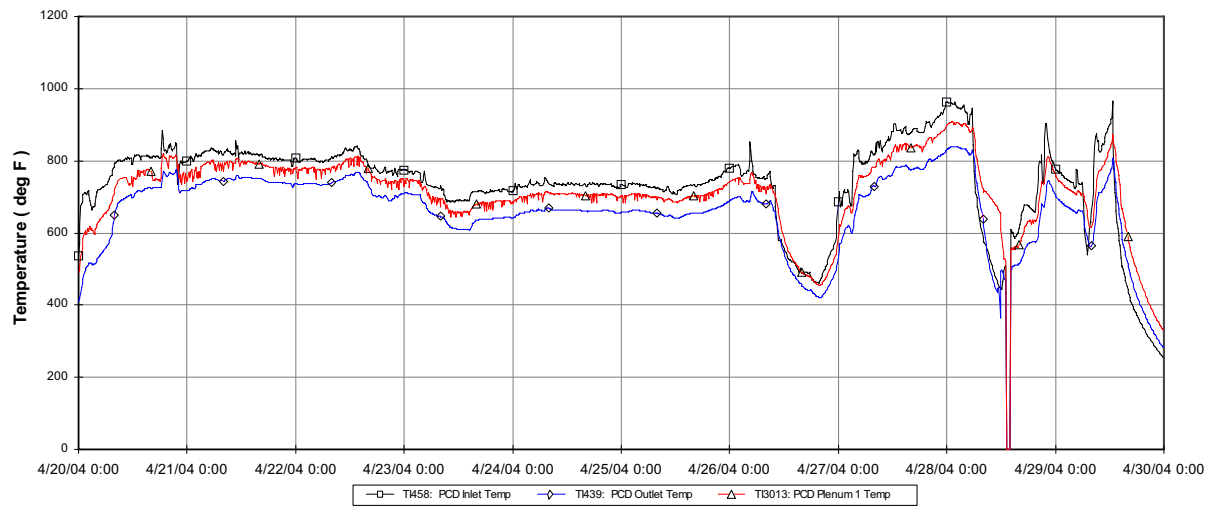


Figure A4-12 PCD Temperatures

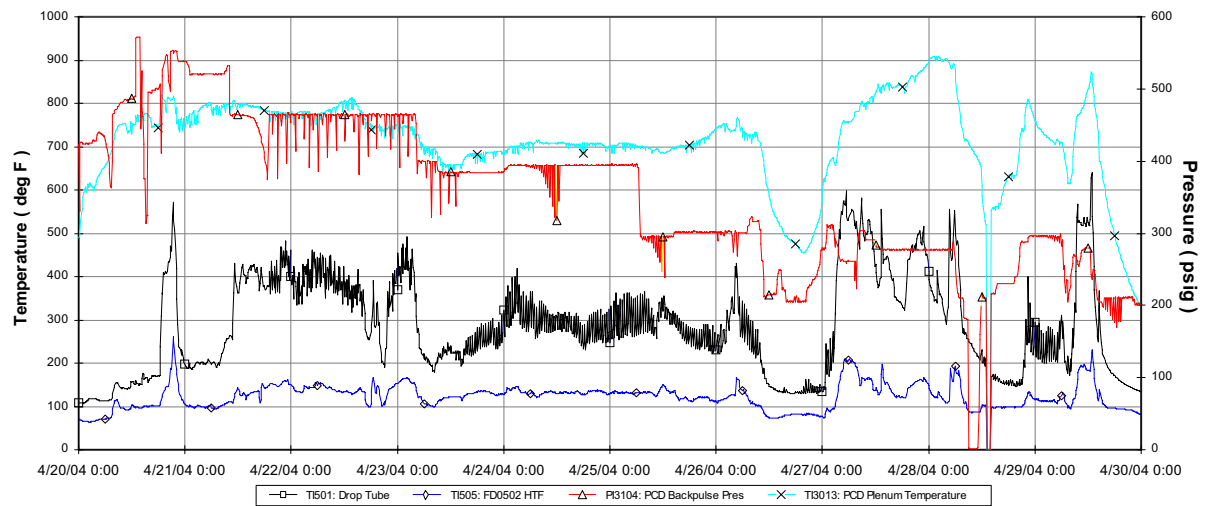


Figure A4-13 PCD Ash Temperatures

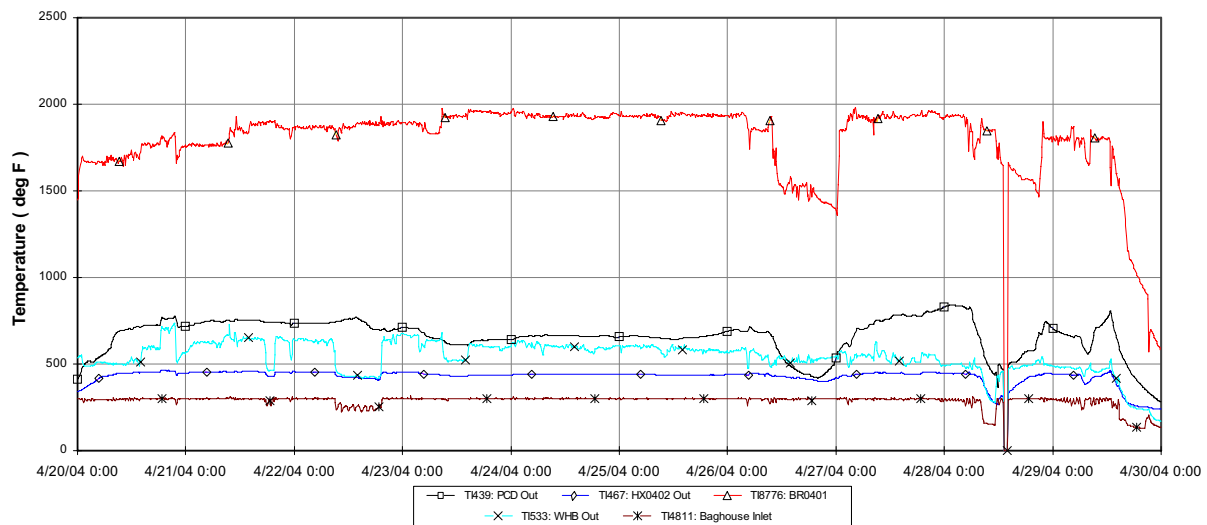


Figure A4-14 System Temperatures Downstream of PCD

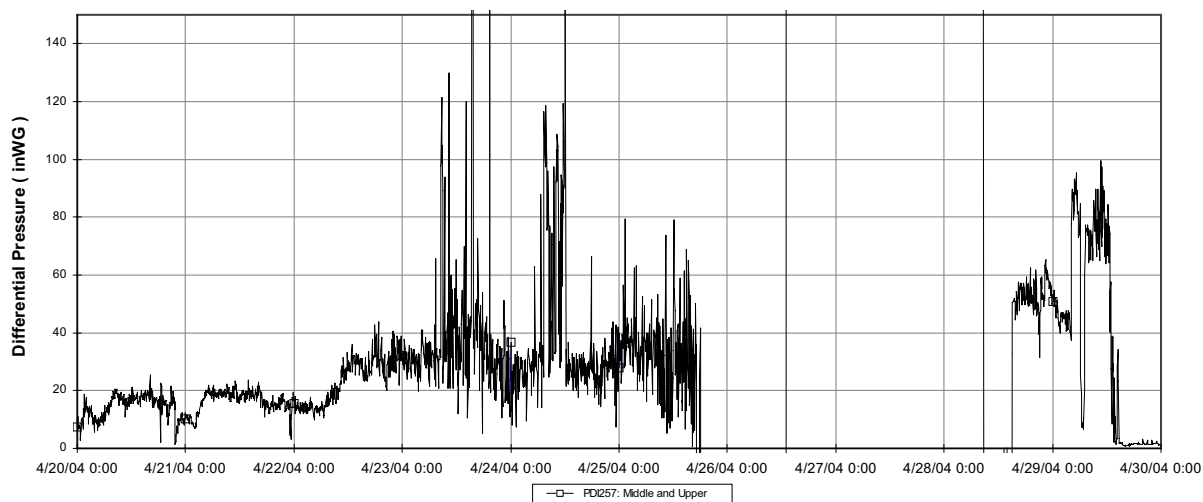


Figure A4-15 Mixing Zone Differential Pressure Profile

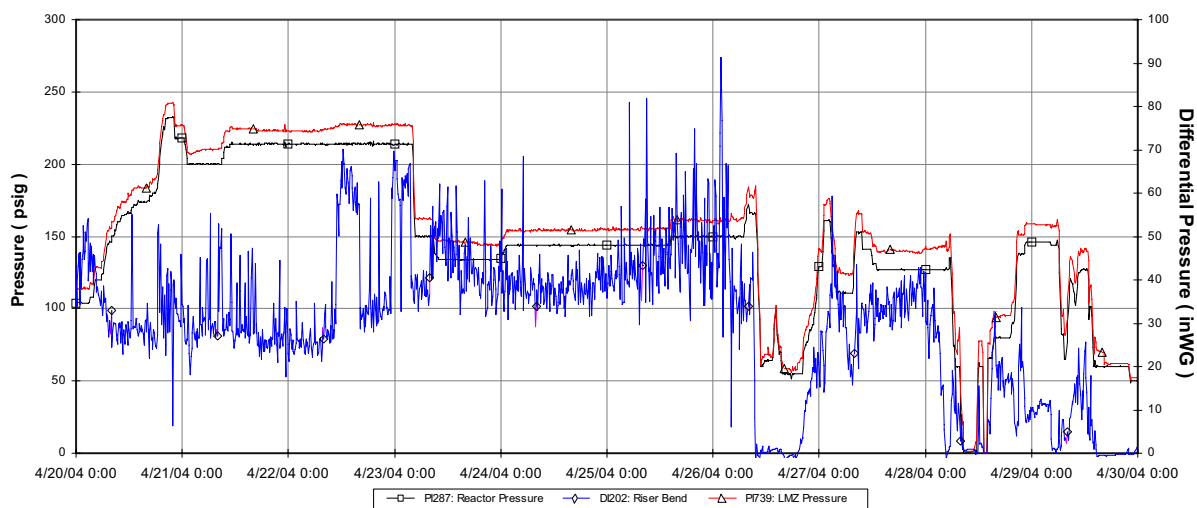


Figure A4-16 Gasifier Pressure/Riser Differential Pressure Profiles

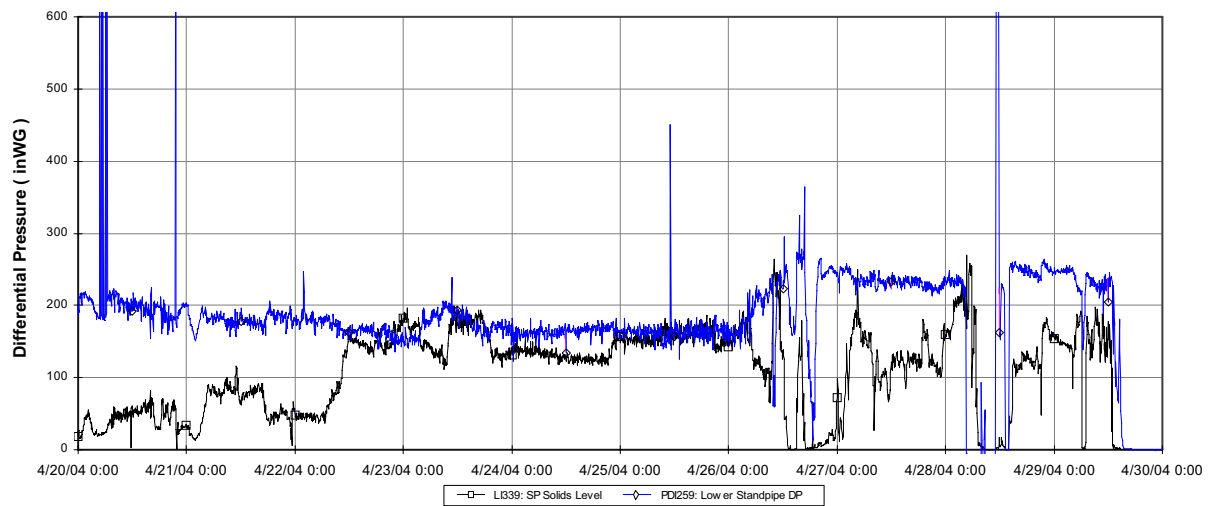


Figure A4-17 Standpipe Differential Pressure Profiles

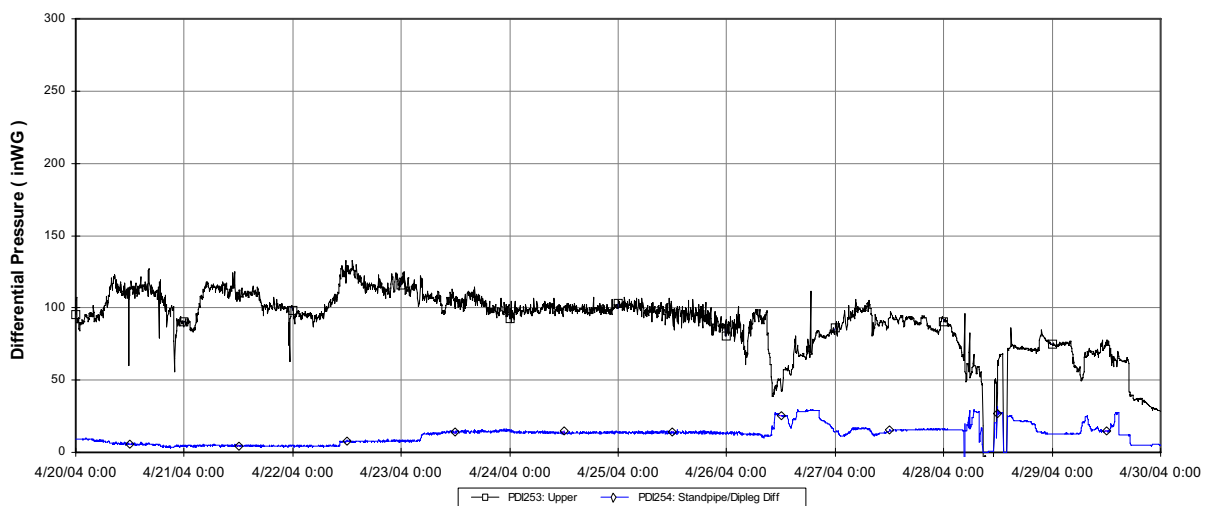


Figure A4-18 Loop Seal Differential Pressure Profiles

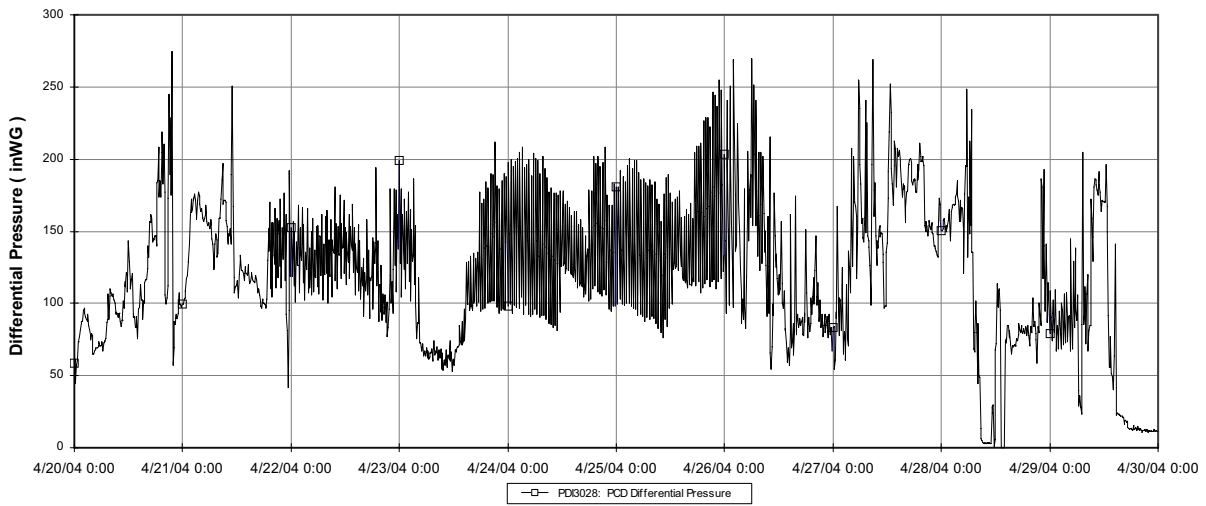


Figure A4-19 PCD Differential Pressure Profile

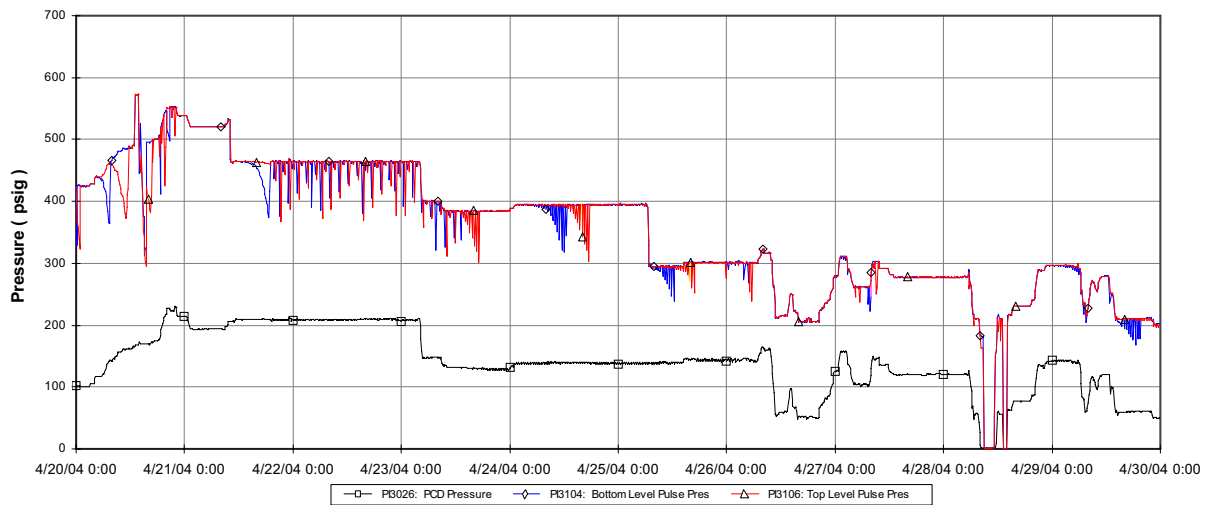


Figure A4-20 PCD Pressure and Pulse Pressure

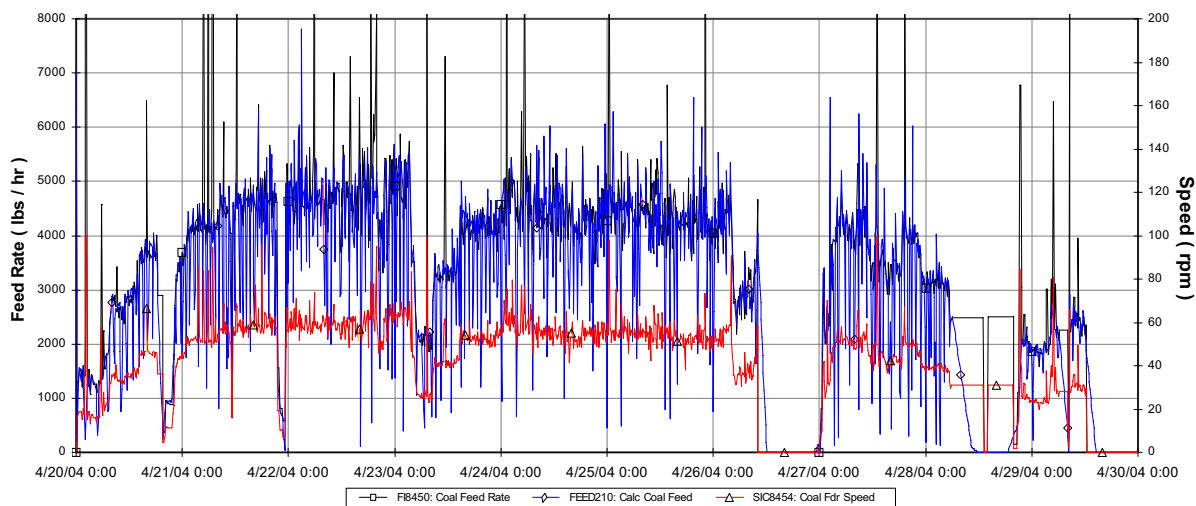


Figure A4-21 Coal-Feed Rate

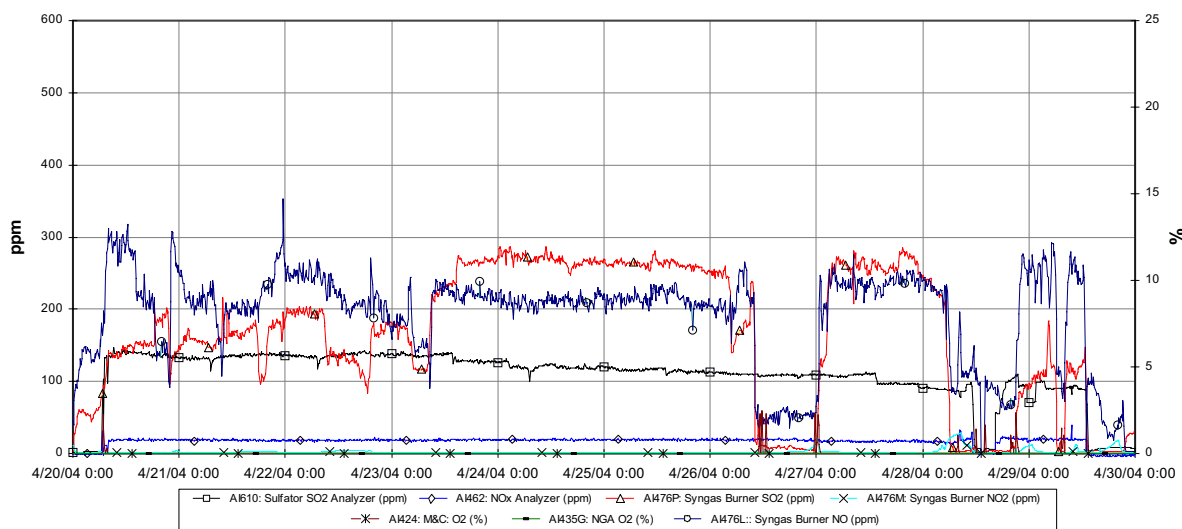


Figure A4-22 O₂, SO₂, and NO_x Analyzers

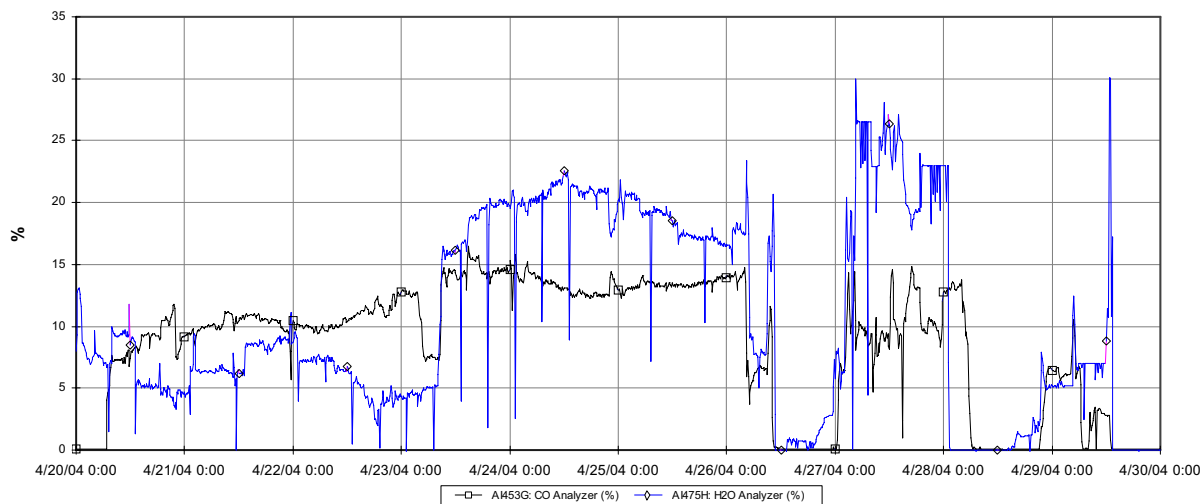


Figure A4-23 CO and H₂O Analyzers

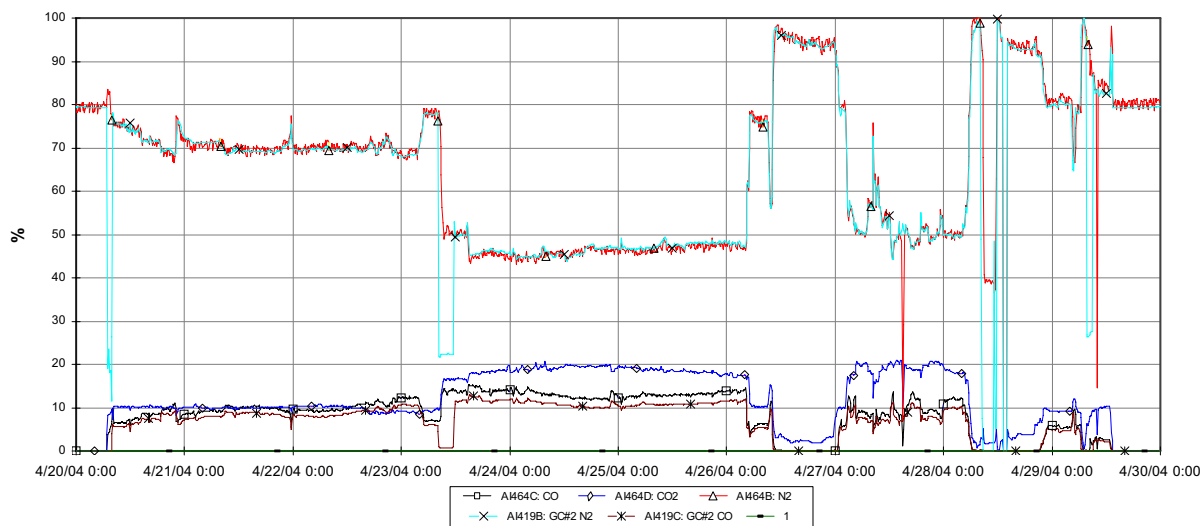


Figure A4-24 Other GC Analysis

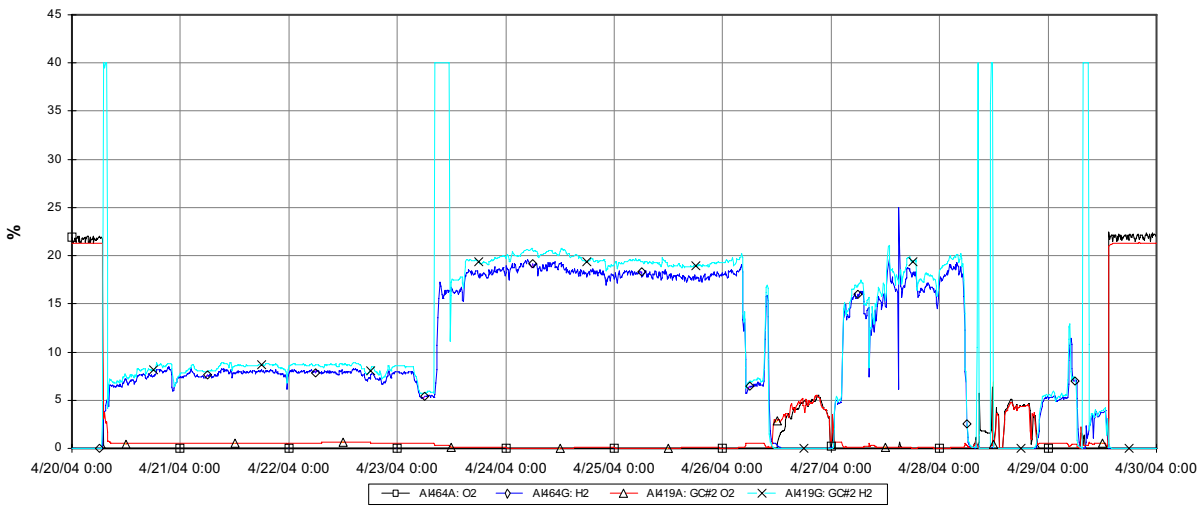


Figure A4-25 O₂ and H₂ Analyzers

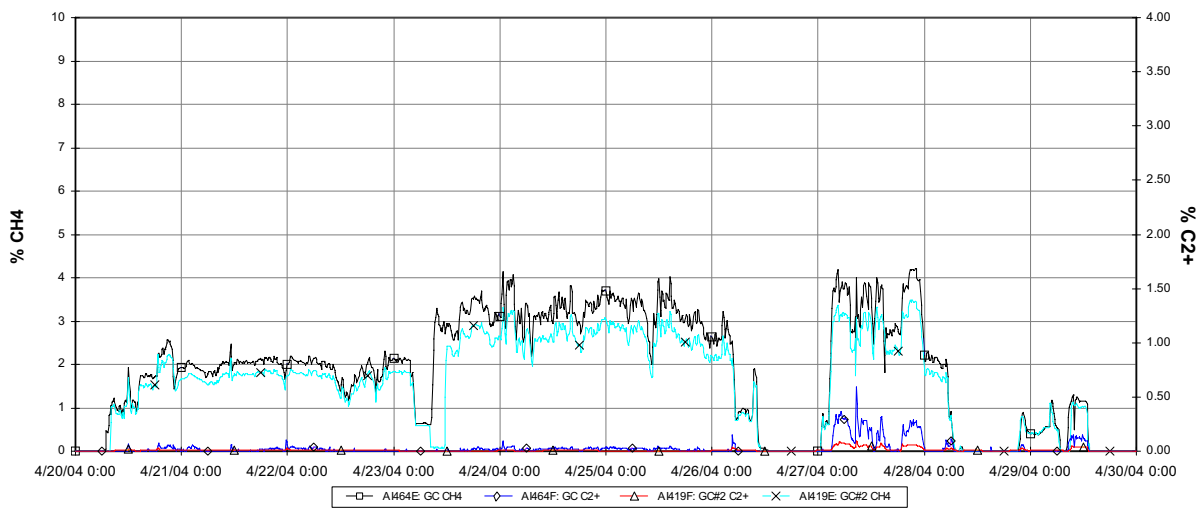


Figure A4-26 C₂⁺ and CH₄ Analyzers

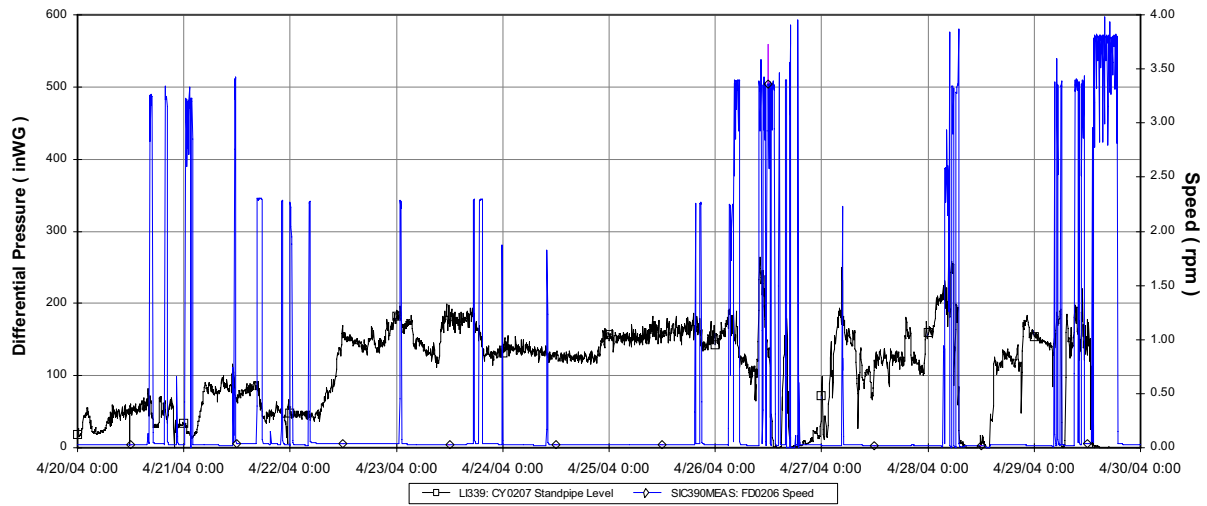


Figure A4-27 Solids Withdrawal From Gasifier

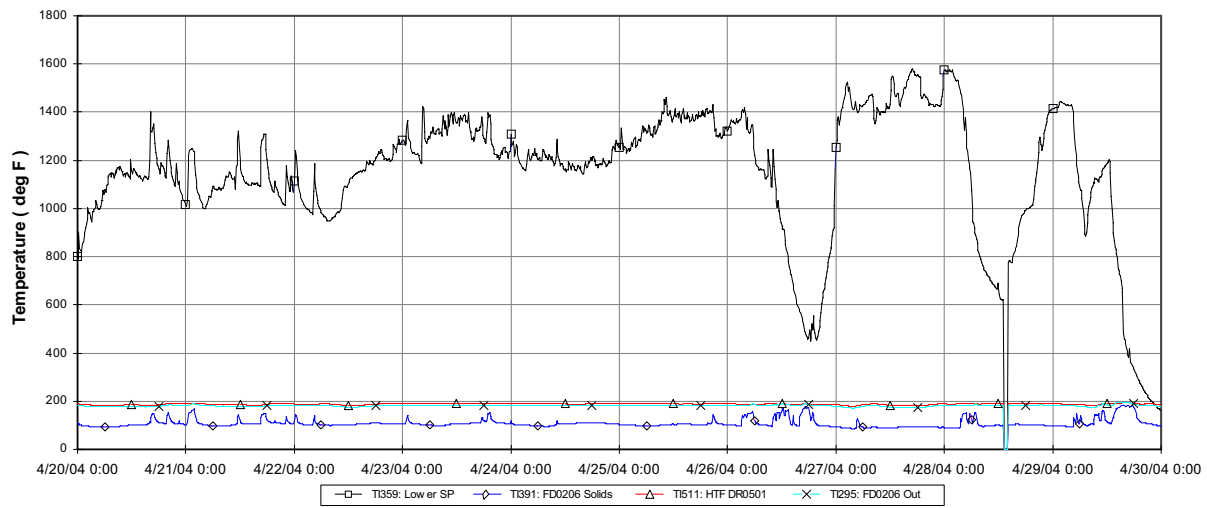


Figure A4-28 FD0510 Temperature Profiles

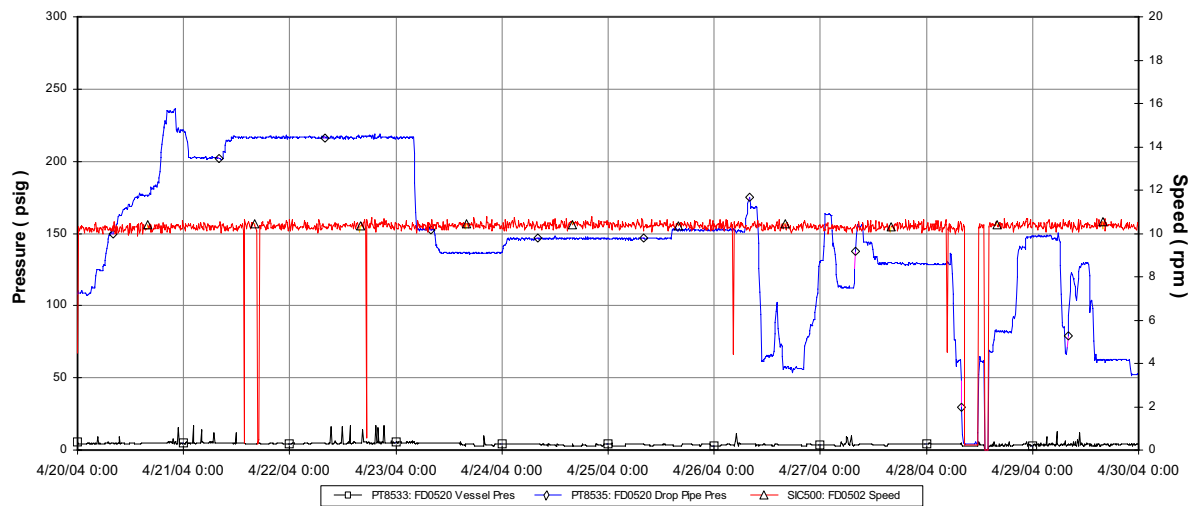


Figure A4-29 FD0520 Pressure Profiles

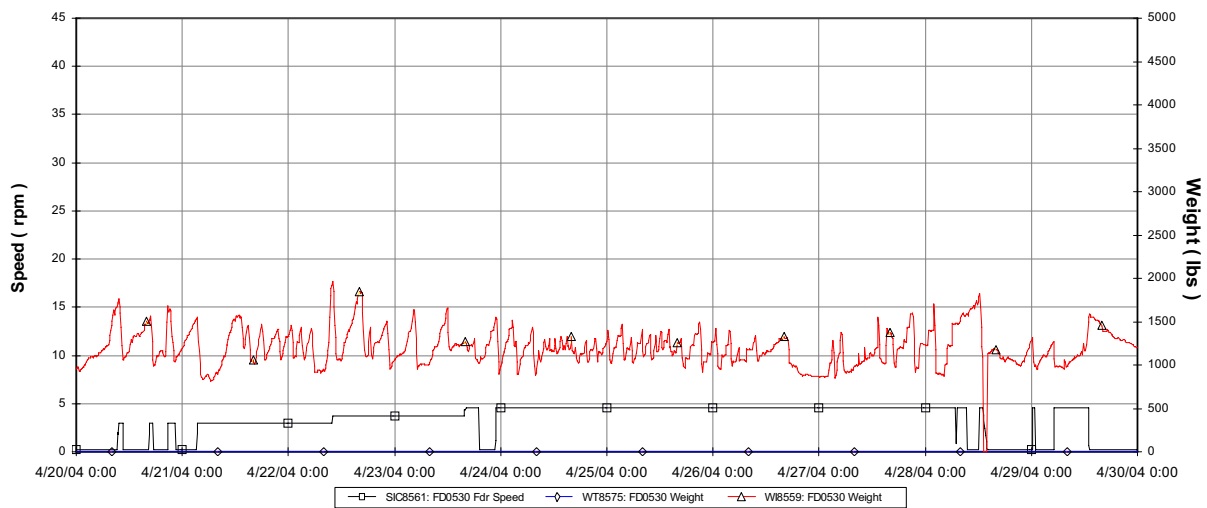


Figure A4-30 FD0530 AFBC Feeder

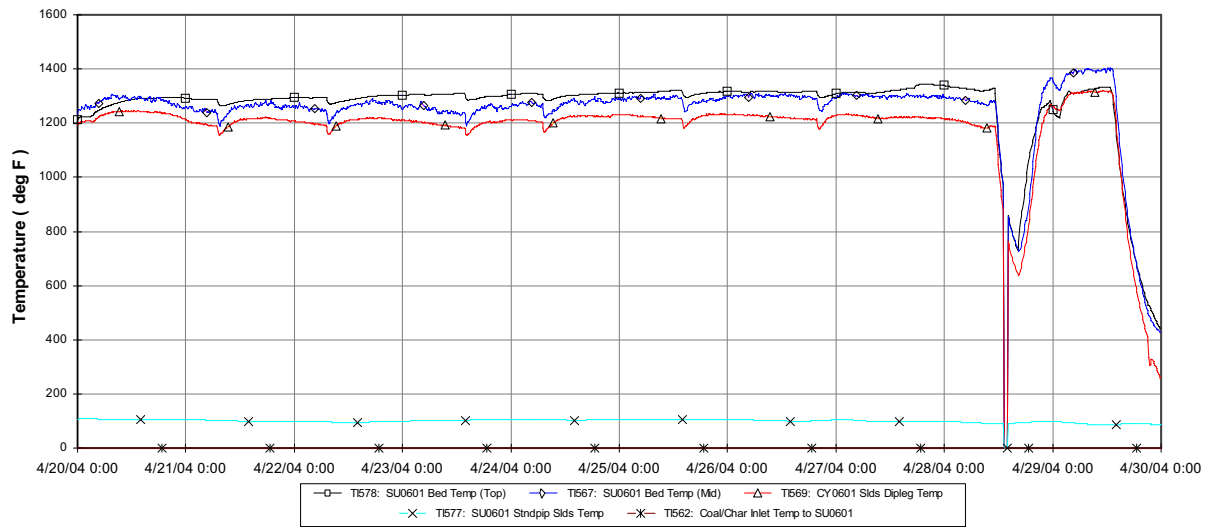


Figure A4-31 AFBC Temperature Profile

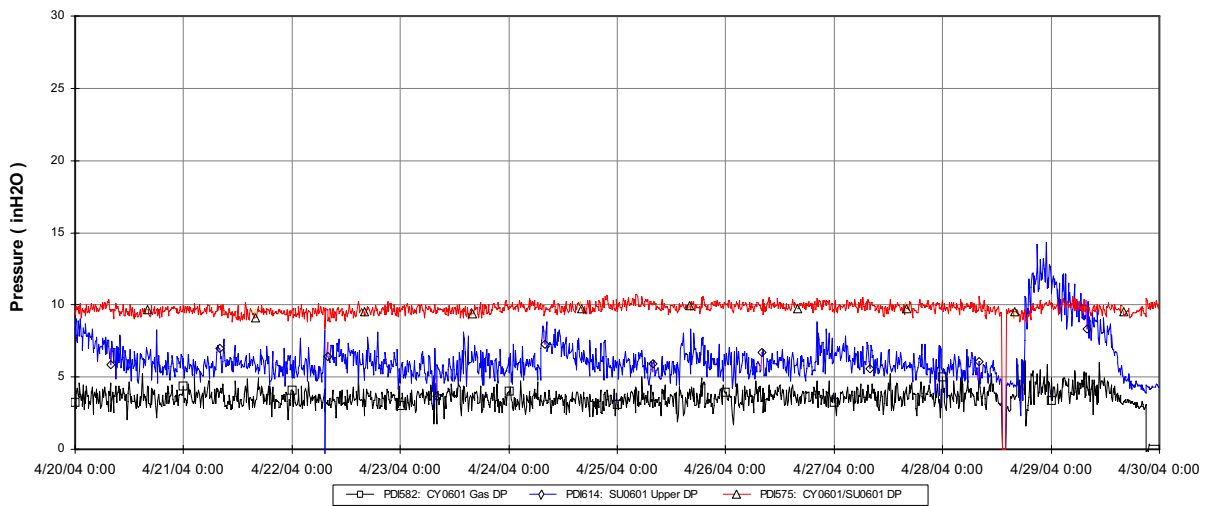


Figure A4-32 AFBC Differential Pressure Profile

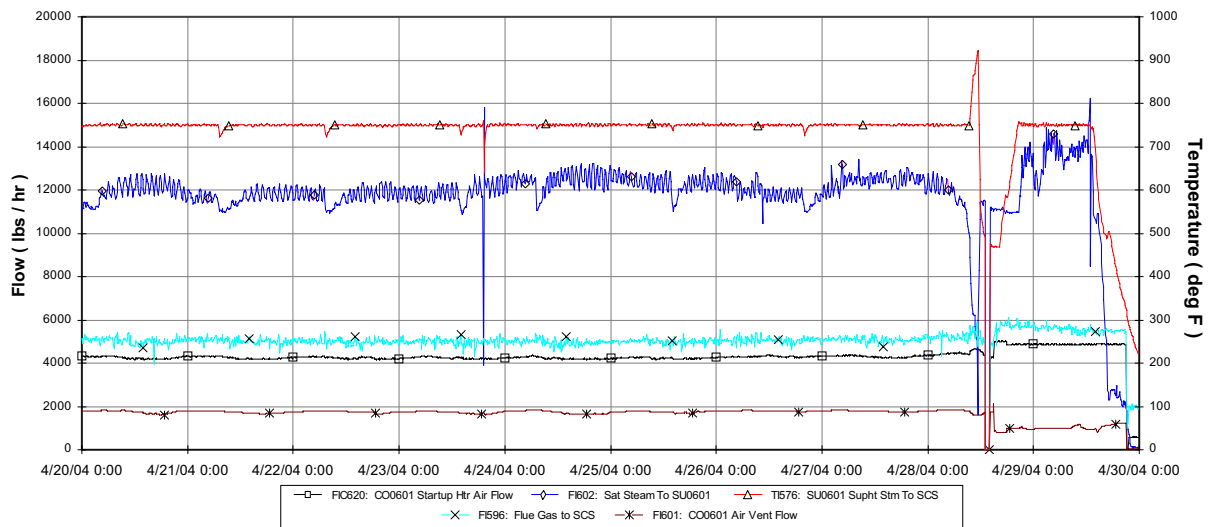


Figure A4-33 AFBC Air/Steam Flows

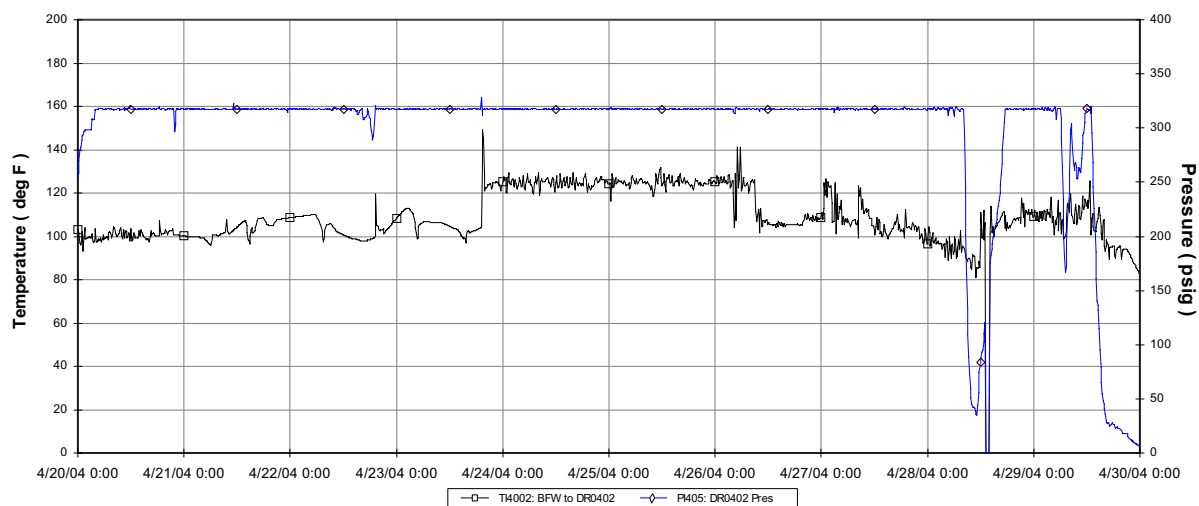


Figure A4-34 Steam Drum Pressure and BFW Temperatures

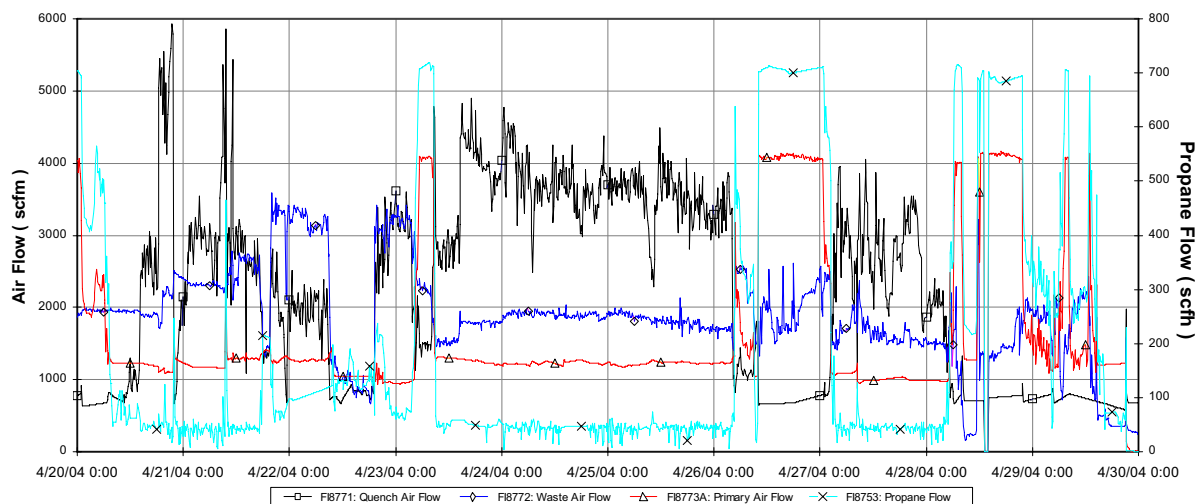


Figure A4-35 Atmospheric Syngas Combustor Air/Fuel Flow

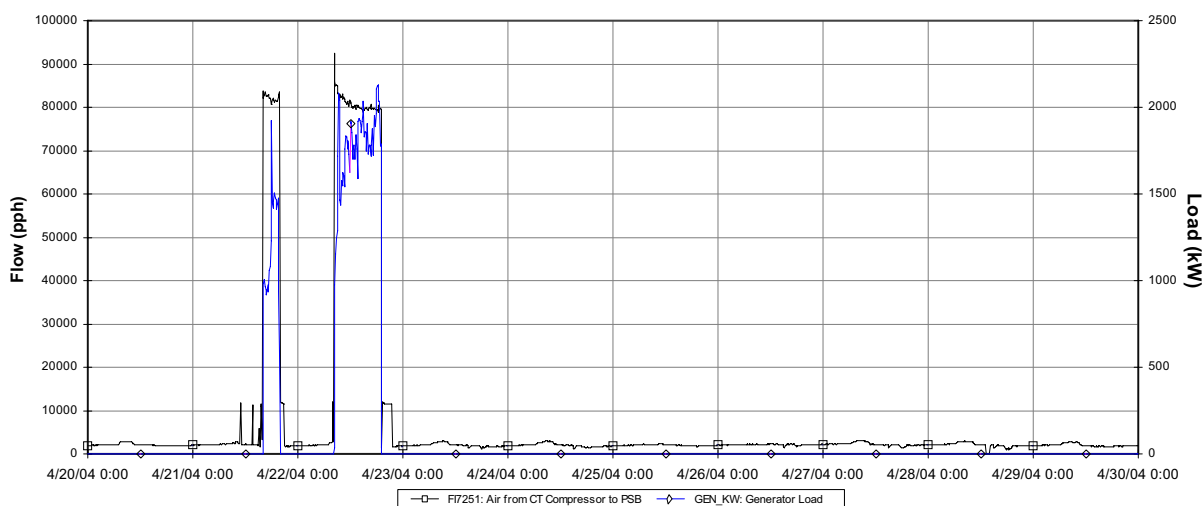


Figure A4-36 Turbine Loading and Air Flow

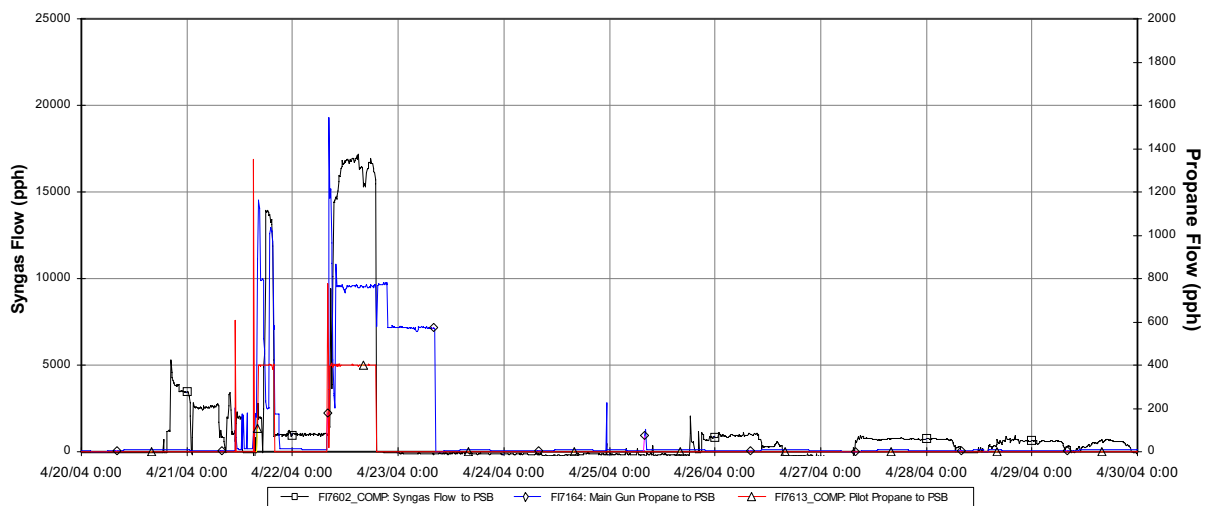


Figure A4-37 Syngas/Propane to PSB

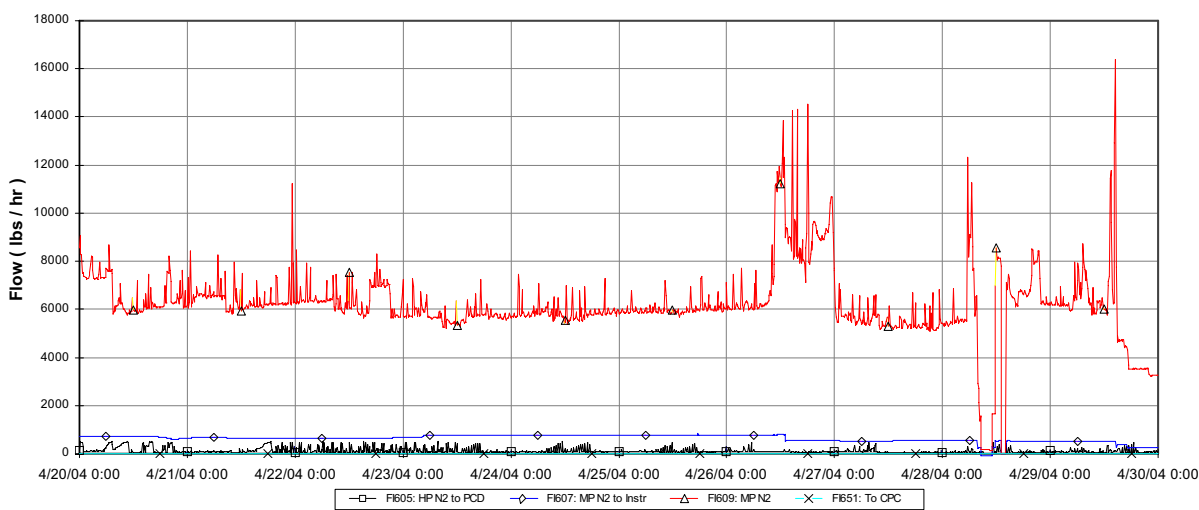


Figure A4-38 Nitrogen Flow Rates

APPENDIX A5 LHV PROJECTION CALCULATIONS

To project a commercial syngas LHV, the following adjustments are made to the raw syngas composition:

1. All non-air nitrogen will be removed from the syngas. A commercial plant will have substantially less instrumentation than the PSDF. Because each individual instrument in a commercial plant will require the same purge-flow rate as the corresponding instrument at the PSDF, the total instrument purge-flow rate will be less. It is assumed that recycled syngas will be used in a commercial plant for aeration. This correction has the effect of increasing all the non-nitrogen syngas compositions and decreasing the nitrogen syngas composition. The recycle syngas flow enters the compressor after the “cold” gas cleanup system. Since the total amount of nitrogen entering the system is reduced, less coal energy will be required to heat the nitrogen and the coal and air/oxygen feed rates will decrease accordingly. It is assumed that this coal would have been combusted to CO_2 and H_2O . Eliminating this additional coal reduces the syngas CO_2 and H_2O concentrations. The lower projected air rates for air-blown mode also decrease the nitrogen content in the projected syngas, and thus decreases the syngas flow rate. The CO/CO_2 ratio will change due to the reduction in CO_2 . This calculation requires an estimated recycle gas-flow rate and an estimated steam-aeration rate to determine the heat required to heat the recycle gas to system temperature. The recycle gas-flow rate is estimated to be 2.4 percent of the syngas-flow rate from the gasifier and is available at 235°F. The aeration steam-flow rate is estimated to be 1.45 percent of the syngas-flow rate from the gasifier and available at 660°F.
2. Small-scale pilot and demonstration units, such as the PSDF, have higher surface area-to-volume ratios than their scaled-up commercial counterparts. Since the heat loss of a commercial plant is difficult to estimate, the projected heat loss is assumed to be zero (adiabatic). The coal, air, and oxygen rates are reduced; the syngas CO_2 , H_2O , and N_2 concentrations are reduced; the CO/CO_2 ratio change. (Based on energy balance data.) The heat loss for the PSDF Transport gasifier is approximately 3.5 MBtu/hr.
3. The steam-flow rate will be adjusted. The steam to oxygen ratio will be the same for the PSDF and the commercial Transport Gasifier. Since steps 1 and 2 reduce the amount of oxygen required, the steam flow rate will decrease correspondingly. The effect of lowering the steam rate will decrease the amount of H_2O in the syngas by the amount the steam rate was reduced. The steam rate and the H_2O content of the syngas are reduced, and hence, the LHV also changes.
4. The water-gas shift is recalculated to reflect the gasifier exit temperature. Corrections 2, 3, and 4 all change the water-gas shift equilibrium constant without changing the gasifier exit temperature. The commercial plant will operate at the PSDF gasifier exit temperature, and hence, have the same water-gas shift equilibrium constant. The H_2O , CO_2 , CO , and H_2 concentrations are then adjusted water-gas shift equilibrium for the temperature of that particular operating period. The LHV could increase if H_2 and CO_2 are converted to H_2O and CO , since the LHV for CO is higher than H_2 . The LHV will decrease if H_2O and CO are converted to H_2 and CO_2 . The LHV correction is usually small, but the change in composition is important if the syngas is used in a fuel cell or for chemical production where the H_2 concentration is a critical design parameter.

5. The commercial plant will use a cold syngas cleanup train that will drop the syngas temperature to 150°F before being reheated prior to entering the gas turbine. At these conditions, moisture will condense from the syngas and exit via a liquid stream. For the commercial design at 388 psia, the syngas water composition at the gas turbine inlet is 0.96 percent. Thus, the final step reduces the syngas moisture content to this value and adjusts the other contents accordingly.

The result of all of these corrections is the commercially projected LHV. Changes 1 and 2 will increase the oxygen-blown LHV more than for the air-blown LHV. One-Hundred percent of the syngas nitrogen will be removed in the oxygen-blown projection, while only about 50 percent of the syngas nitrogen will be removed for the air-blown projection.

These calculations are an oversimplification of the gasification process. A more sophisticated model is required to precisely predict the effects of decreasing pure nitrogen and gasifier heat loss. Note that the projected syngas compositions are based on a projected coal rate, projected air rate, projected oxygen rate, projected steam rate, and a projected syngas rate.

APPENDIX A6 TERMS

Listing of Abbreviations

AFBC	Atmospheric Fluid-Bed Combustor
CFAD	Continuous Fine Ash Depressurization
CT	Combustion Turbine
DOE	U.S. Department of Energy
EPRI	Electric Power Research Institute
FTIR	Fourier Transform Infrared Spectroscopy Analyzer
FPD	Flame Photometric Detector
FV	Face Velocity
GC	Gas Chromatograph
HHV	Higher Heating Value
IGCC	Integrated Gasification Combined Cycle
KBR	Kellogg, Brown and Root, Inc.
LHV	Lower Heating Value
LMZ	Lower Mixing Zone
LOI	Loss on Ignition
M%	Molar Percent
MMD	Mass Mean Diameter
NCC	Non-Carbonate Carbon
NM	Not Measured
PCD	Particulate Control Device
PCME	Pollution Control and Measurement (Europe)
PRB	Powder River Basin
PSB	Piloted Syngas Burner
PSDF	Power Systems Development Facility
RAPTOR	Resuspended Ash Permeability Tester
RT	Room Temperature
SMD	Sauter Mean Diameter
SPG	Siemens Power Generation
TRS	Total Reduced Sulfur
UMZ	Upper Mixing Zone

Listing of Units

acfm	actual cubic feet per minute
Btu	British thermal units
°C	degrees Celsius or centigrade
°F	degrees Fahrenheit
ft	feet
FPS	feet per second
gpm	gallon per minute
g/cm ³ or g/cc	grams per cubic centimeter
g	grams
GPa	gigapascals
hp	horsepower
hr	hour
in	inches
inH ₂ O	inches water
inWg (or in Wc)	inches, water guage (inches, water column)
in-lb	inch pounds
°K	degrees Kelvin
kg	kilograms
kJ	kilojoules
kPa	kilopascals
ksi	thousand pounds per square inch
m	meters
min	minute
mm	millimeters
MPa	megapascals
Msi	million pounds per square inch
MW	megawatts
m/s	meters per second
MBtu	million British thermal units
m ² /g	square meters per gram
μ or μm	microns or micrometers
dp ₅₀	particle size distribution at 50 percentile
ppm	parts per million
ppm(v)	parts per million (volume)
ppm(w)	parts per million (weight)
lb	pounds
pph	pounds per hour
psi	pounds per square inch
psia	pounds per square inch absolute
psid	pounds per square inch differential
psig	pounds per square inch gauge
ΔP	pressure drop
rpm	revolutions per minute
s or sec	seconds
scf	standard cubic feet

scfh	standard cubic feet per hour
scfm	standard cubic feet per minute
V	volts
W	watts